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COMMERCIAL DEMONSTRATION OF THE NOXSO SO₂/NO_x REMOVAL FLUE GAS CLEANUP SYSTEM

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EXECUTIVE SUMMARY

The NOXSO process is a dry, post-combustion flue gas treatment technology which uses a regenerable sorbent to simultaneously adsorb sulfur dioxide (SO_2) and nitrogen oxides (NO_x) from the flue gas of a coal-fired utility boiler. In the process, the SO_2 is reduced to sulfur by-product (elemental sulfur, sulfuric acid, or liquid SO_2) and the NO_x is reduced to nitrogen and oxygen. It is predicted that the process can economically remove 90% of the acid rain precursor gases from the flue gas stream in a retrofit or new facility.

The objective of the NOXSO Demonstration Project is to design, construct, and operate a flue gas treatment system utilizing the NOXSO process at Ohio Edison's Niles Plant Unit #1. The effectiveness of the process will be demonstrated by achieving significant reductions in emissions of sulfur and nitrogen oxides. In addition, sufficient operating data will be obtained to confirm the process economics and provide a basis to guarantee performance on a commercial scale.

The project is presently in the project definition and preliminary design phase. Data obtained during pilot plant testing which was completed on July 30, 1993 is being incorporated in the design of the commercial size plant. A suitable host site to demonstrate the NOXSO process on a commercial scale is presently being sought.

Preliminary engineering activities focused on an alternate general arrangement featuring a low profile design. This design features obround, partially shop fabricated vessels and high temperature dense phase transport systems. The incentive for this design is to reduce the capital cost and construction duration. Cost reduction should be realized due to less costly shop fabrication of vessels, lower plant height reducing the quantity of structural steel, and lower plant height reducing the complexity of the foundation system. This design introduces an additional technical risk since the high temperature dense phase transport system was not tested at the pilot plant. Qualified vendors to provide a plant for making elemental sulfur from the regenerator offgas stream have been identified.

Testing at the pilot plant was completed on July 30, 1993. The pilot plant results were better than originally anticipated with regard to all the important measures of performance including removal efficiency, energy consumption, and sorbent attrition rate. Modifications to the adsorber to include a second adsorber bed and cooling water sprays resulted in SO_2 and NO_x removal efficiencies of 99% and 88% respectively. These removal efficiencies were obtained with inlet SO_2 and NO_x concentrations of 2650 ppm and 971 ppm respectively.

Specific activities conducted to support the demonstration plant design include: an updated adsorber model, computer process simulator, sulfur by-product market and technical study, HCl removal by the process, and L-valve design development.

1.0 INTRODUCTION

The NOXSO process is a dry, post-combustion flue gas treatment technology which uses a regenerable sorbent to simultaneously adsorb sulfur dioxide (SO_2) and nitrogen oxides (NO_x) from the flue gas of a coal-fired utility boiler. In the process, the SO_2 is converted to a sulfur by-product (elemental sulfur, sulfuric acid, or liquid SO_2) and the NO_x is reduced to nitrogen and oxygen. It is predicted that the process can economically remove 90% of the acid rain precursor gases from the flue gas stream in a retrofit or new facility.

Details of the NOXSO process are described with the aid of **Figure 1-1**. Flue gas from the power plant is drawn through a flue gas booster fan which forces the air through a two-stage fluid bed adsorber and centrifugal separator before passing to the power plant stack. Water is sprayed directly into one or both of the fluid beds as required to lower the temperature to 250-275°F by evaporative cooling. The fluid bed adsorber contains active NOXSO sorbent. The NOXSO sorbent is a 1.6 mm diameter γ -alumina bead impregnated with 5.2 weight % sodium. The centrifugal separator separates sorbent which may be entrained in the flue gas and returns it to the inlet of the dense phase transport system.

Spent sorbent from the adsorber flows into a dense-phase conveying system which lifts the sorbent to the top bed of the sorbent heater vessel. The sorbent flows through the multi-stage fluidized bed sorbent heater counter to the heating gas which heats the sorbent to the regeneration temperature of approximately 1150°F.

In the process of heating the sorbent, the NO_x is driven from the sorbent and carried to the power plant boiler in the NO_x recycle stream. The NO_x recycle stream is cooled from approximately 500°F to 150°F in the feedwater heater. This heater heats a slip stream of the power plants feedwater, thereby reducing the amount of extraction steam taken from the low pressure turbine, enabling the generation of additional electricity. The cooled NO_x recycle stream replaces a portion of the combustion air. The presence of NO_x in the combustion air suppresses the formation of NO_x in the boiler resulting in a net destruction of NO_x .

The heated sorbent enters the regenerator where it is contacted with natural gas. Through a series of chemical reactions, the sulfur on the sorbent combines with the methane and forms SO_2 and H_2S . Additional regeneration occurs in the steam treater when the sorbent is contacted with steam converting the remaining sulfur on the sorbent to H_2S .

The regenerator and steam treater offgas streams are combined and directed to a sulfur recovery plant where the H_2S and SO_2 are converted to a sulfur by-product. Elemental sulfur, sulfuric acid, and liquid SO_2 are all potential end products from the regenerator offgas stream. Tail gas from the sulfur recovery plant will be incinerated and recycled back through the adsorbers to remove any sulfur compounds.

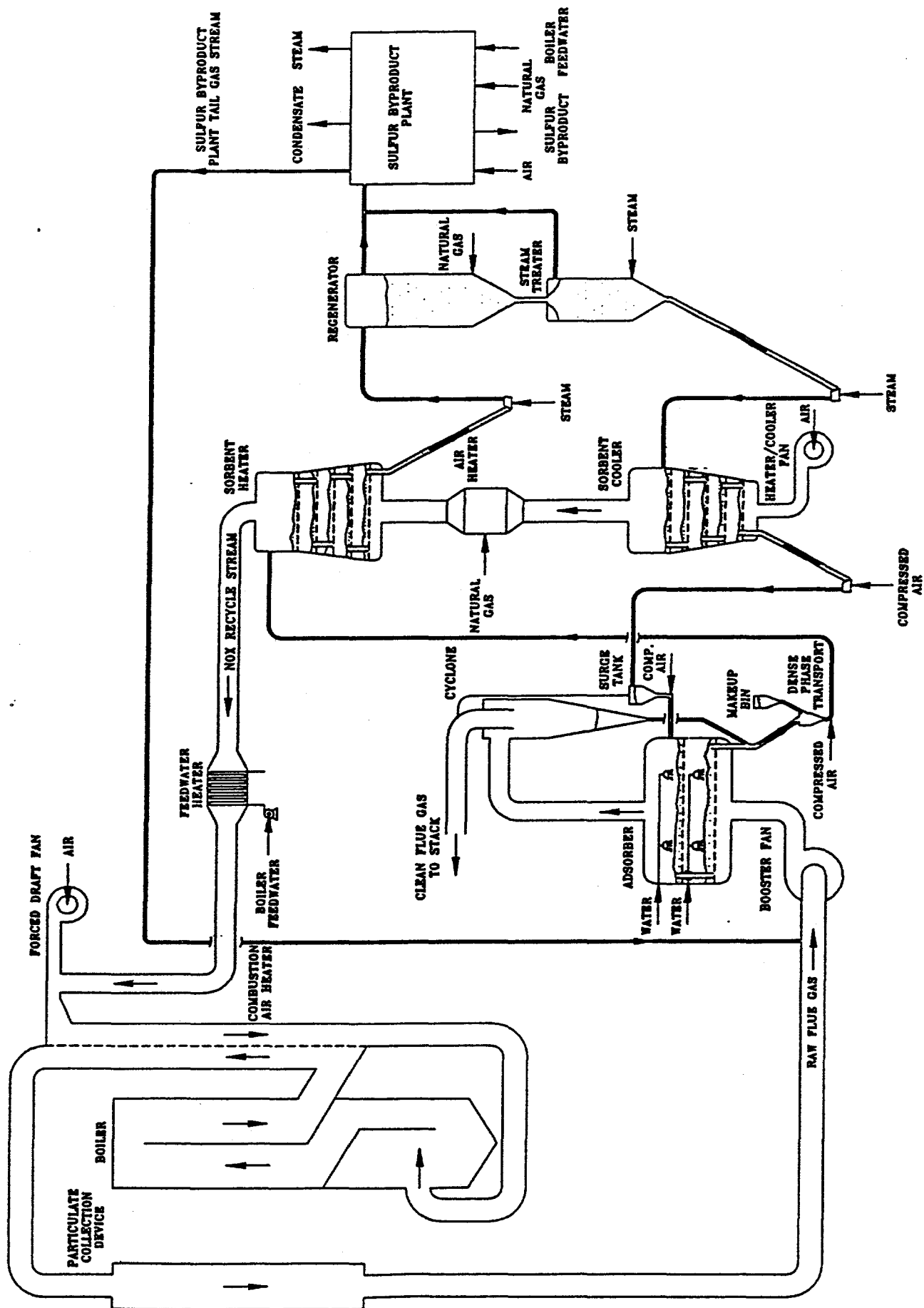


Figure 1-1. NOXSO Process Diagram

High temperature sorbent exiting the steam treater passes to the multi-stage fluidized bed sorbent cooler. The sorbent flows counter to the ambient air which cools the sorbent. Regenerated sorbent exits the cooler at 300°F. It is directed to the adsorber completing the sorbent cycle.

Ambient air which is forced through the sorbent cooler by the heater-cooler fan exits the sorbent cooler at approximately 900°F. This preheated air then enters the air heater where it is heated to approximately 1350°F so it is capable of heating the sorbent exiting the sorbent heater to 1150°F.

2.0 PROJECT DESCRIPTION

The objective of the NOXSO Demonstration Project is to design, construct, and operate a commercial scale flue gas treatment system utilizing the NOXSO process. The effectiveness of the process will be demonstrated by achieving significant reductions in emissions of sulfur and nitrogen oxides. In addition, sufficient operating data will be obtained to confirm the process economics and provide a basis to guarantee performance on a commercial scale.

3.0 PROJECT STATUS

The project is presently in the project definition and preliminary design phase. This phase was included in the project to allow completion of studies and preliminary activities which could be conducted in parallel with NOXSO's pilot plant project being conducted at Ohio Edison's Toronto Power Plant.

Testing at the pilot plant was completed on July 30, 1993. Performance at the pilot plant exceeded the design expectations for pollutant removal efficiency, sorbent attrition, and power and natural gas consumption. The data continues to be evaluated in the form of developing design equations which can be used in the design of the commercial plant.

Installation of the demonstration plant was originally planned for Ohio Edison's Niles Plant Unit #1. Due to the inability to negotiate an arrangement for operation of the NOXSO plant after the demonstration, a new host site is being sought.

3.1 Project Management

Project management activities included the routine efforts of task assignments and information transfer within the project team comprised of personnel from NOXSO, MK-Ferguson, and W.R. Grace. Additionally, all DOE reports were submitted in accordance with the reporting requirements of the cooperative agreement. Significant effort was expended negotiating agreements necessary for the novation of the cooperative agreement from MK-Ferguson to NOXSO.

3.2 NEPA Compliance

A draft EIV and first draft of an environmental assessment (EA) had been completed based on the demonstration being conducted at Ohio Edison's Niles Power Plant. A significant portion of these documents is specific to the location of the project. Consequently, the NEPA process is on hold until the new host site is identified.

3.3 Preliminary Engineering

During this quarter, information to assess the low profile plant design both technically and economically was developed. A second sulfur plant design and vendor were evaluated. The construction cost estimate for the low profile plant located at a generic site is almost complete.

3.3.1 Low Profile Plant Design

A low profile plant design (**Figure 3-1**) is being developed as an alternative to the scaled up pilot plant design (**Figure 3-2**) in an effort to reduce the capital cost and construction duration. This low profile design is approximately 100 feet shorter than the original design while having substantially the same footprint.

The major features of this design are horizontal vessels for the adsorber, sorbent heater and sorbent cooler, and the use of dense phase solid transport systems in place of the J-valves. The horizontal vessels are of obround cross-section. They will be shop fabricated in three sections, then shipped to the site where the grids will be installed. The sections will be joined and the vessels will be raised into position. Each vessel will be an upper and lower section of six foot radius and a center section eight feet high. Each vessel will contain two grids. On site construction time will be reduced and the size and reach of any crane necessary to install the vessels or erect the structure will be minimized.

Dense phase solid transport systems have been specified for use due to the higher sorbent lifts required by this design. The use of an L-valve is being considered for this service.

This design will result in less complicated foundations and structural steel. Piping, ductwork, and electrical tray and conduit quantities will be reduced. In addition to the advantages the low profile design provides for the demonstration plant, it is a fully modular design adaptable to boilers of different capacity by varying the length of the process vessels or adding modules as required.

3.3.1.1 P&ID's

The low profile plant design combines several process design features in a single new design. It is possible to produce intermediate designs which include various permutations and combinations of these design features.

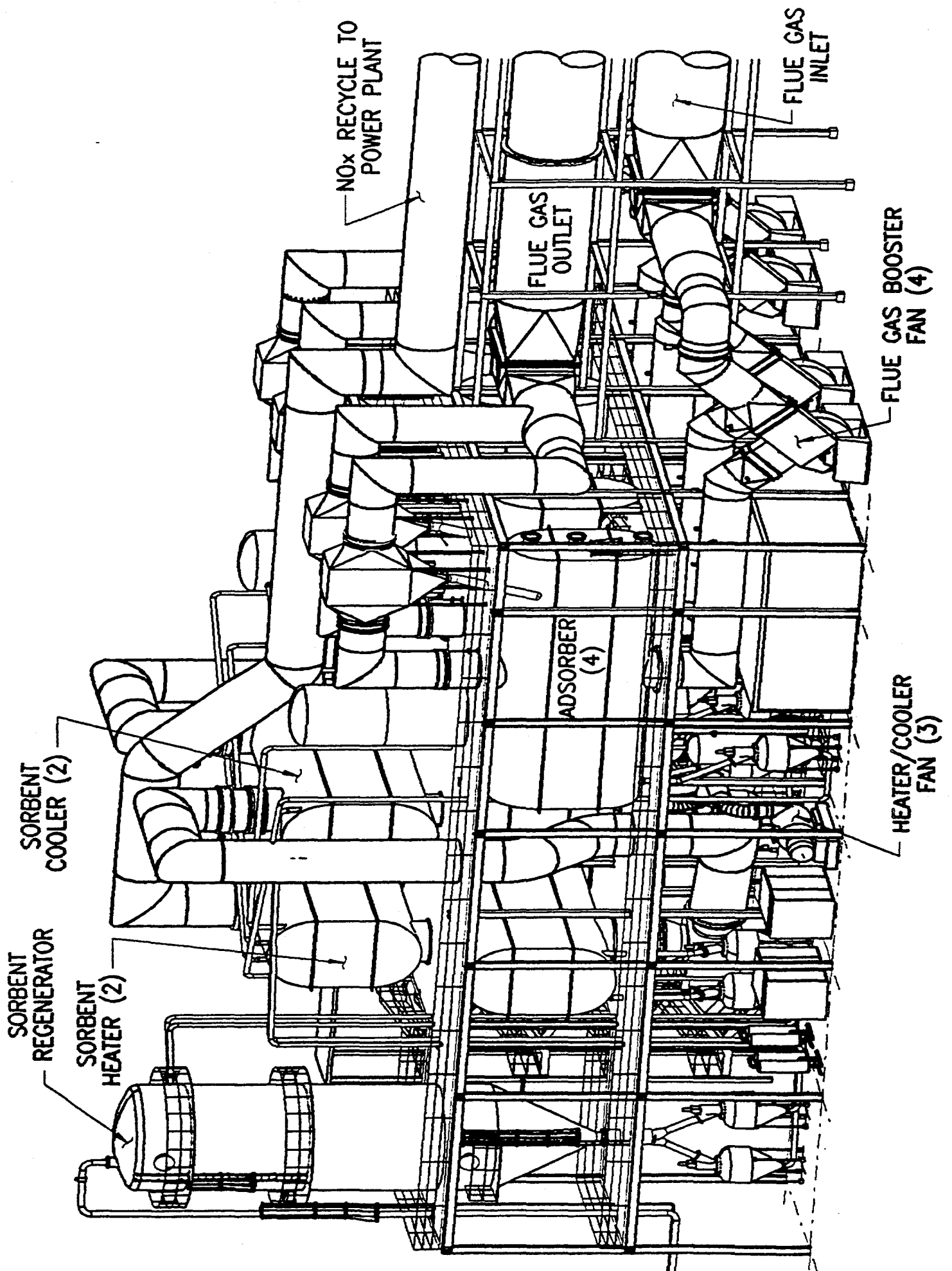


Figure 3-1. Low Profile Plant

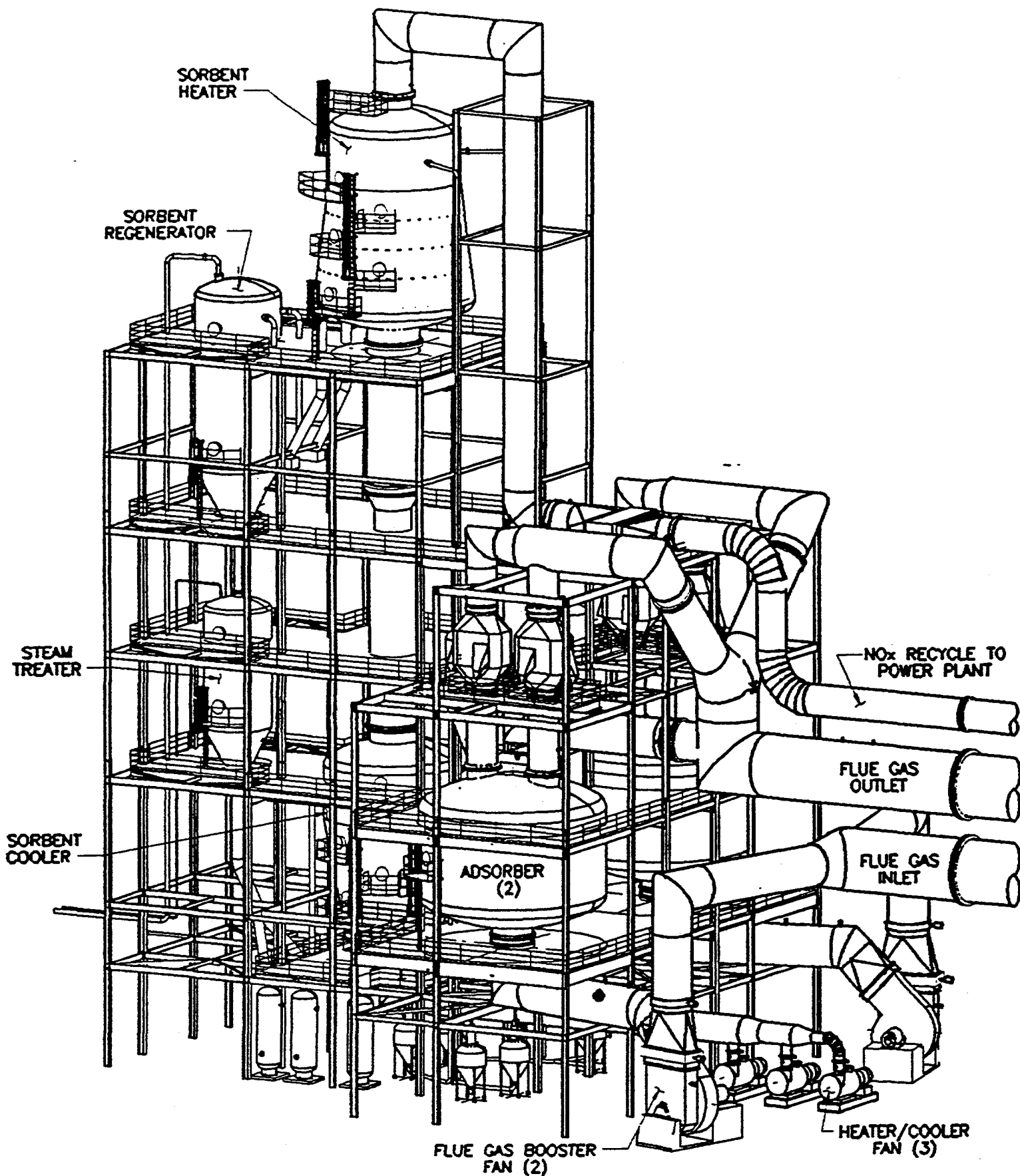


Figure 3-2. NOXSO Process Tower

Shop fabrication of the adsorber, sorbent heater, and sorbent cooler results in long narrow vessels with two fluid bed stages per vessel. The previous adsorber design used two vessels in parallel, each containing two fluid beds. The current design is for four vessels in parallel each with two fluid beds. Four vessels were used to keep vessel lengths reasonable and provide a compact layout. In addition, this design provides better velocity control through the fluid beds during turndown. Four horizontal cyclones are used, the same number as the previous design, but now each adsorber has one dedicated cyclone. There is one adsorber fan for each adsorber to simplify ductwork, valving, flow control, and to allow for the shutdown of one adsorber train. The long narrow vessels result in sorbent plug flow as opposed to the circular vessels where solids are mixed. This may improve the adsorption efficiency of the system; however, some testing will have to be conducted to verify this.

Sorbent heater and cooler systems consist of two vessels containing two fluid beds each for a total of eight fluid beds. Five stages were proposed in the original design for each vessel but shop fabricated vessels are most economical at two stages each. This results in the choice of an even number of stages for the sorbent heater and sorbent cooler. Four stages were chosen for each service (heater and cooler) instead of six stages. Improved heat transfer due to an approach to plug flow by the sorbent was the reason the four-stage design was chosen.

The steam treater was eliminated based on process data obtained from the POC plant. Steam disengagement was added to the top of the regenerator to reduce the amount of steam carry-over to the sulfur plant with the regenerator offgas.

J-valves were replaced with dense phase transport systems. In previous designs the J-valve sorbent transport systems set the relative elevation between the sorbent heater, the regenerator, and the sorbent cooler. The dense phase transport systems eliminated this height requirement and resulted in a shorter plant design. An alternative design using high lift L-valves has been successfully tested at the POC and provides another option for sorbent transport.

3.3.1.2 Systems Control

The instrument list grew by 30% in total number of tags; however, the cost reflected by these totals has not increased. This is mainly due to the J-valve instrumentation being deleted which includes two process analyzers. The increase in tag quantities is attributed to temperature inputs and vendor furnished items on the additional dense phase systems. The sulfur plant instruments are included in the total instrument count.

The distributed control system (DCS) I/O loading was increased and an adjustment was made for this. The I/O list provided by Advanced Petrogas Systems, Inc. for the sulfur recovery plant was added to the DCS requirement. The dense phase system, air heater, and any other vendor supplied equipment will also be controlled from the DCS.

3.3.1.3 Vessels

To attain the highest degree of shop fabrication possible, the process vessels must be constructed in a way that permits transportation over the road. To accomplish this a width of 12 feet was chosen and the length and number of vessels were determined to provide the proper cross-sectional flow area. This resulted in four adsorber vessels in a parallel flow arrangement each 44 feet long. The sorbent cooler required two vessels in a series flow arrangement, one 52 feet long and one 36 feet long. The sorbent heater required two vessels in a series flow arrangement, one 62 feet long and one 45 feet long. The variation in length of the heater and cooler vessels maintains the gas velocity near the nominal design point of 3 ft/sec as temperatures change. All vessels have two grid systems for the fluidized beds.

The cross-section of these vessels is obround with a semi-cylindrical top and bottom and flat sides. The ends of the vessels are flat plates. See **Figure 3-3**. The ASME boiler and pressure vessel code Section VIII, Division 1, Appendix I provides rules and formulas for the design of vessels of this type. This type of construction imposes bending moments on the vessel. The maximum stresses occur at points A, B, and C (**Figure 3-3**). The magnitude of the stresses is dependent on the internal pressure, P , the radius of the semi-cylindrical section, R , and half the length of the flat side L_2 . To prevent stresses from exceeding allowable limits, reinforcement internal stays and external stiffeners can be used. Calculations were made for an externally stiffened adsorber and an internally stayed adsorber vessel. The adsorber calculations are given in Appendix I. The externally stiffened vessel has a wall thickness of 1/2 inch and 8 inch I-beams on a pitch of 2 feet for reinforcement. The internally stayed vessel has 1/2 inch semi-cylindrical sections, 1 inch flat sides and 3/4 inch stays on a pitch of 2 feet. Both configurations have flat end plates 1-1/2 inches thick. The stresses in the externally reinforced vessel were all within allowable limits while those in the internally stayed vessel were not.

A finite element model of an adsorber vessel was programmed using Algor. The model uses external and internal reinforcement in combination. The external stiffeners are 1" x 6" bars and the internal stays are 1" x 12" bars. Both are on a pitch of 5 feet 6 inches. The vessel shell is 5/8 inch thick and the end plates are 1 inch thick. **Figures 3-4** and **3-5** show the stress distribution in this vessel.

The grids are formed from 10 gage and 12 gage stainless steel sheets. Each grid segment is formed into a channel approximately 21 inches wide with 8 inch legs and 1 inch returns on the 8 inch sides. The grid spans the 12 foot dimension of the vessels and is supported on a ledge. Each grid is bolted to adjacent grids and gasketing material is used to form a seal. A clamping angle with slotted bolt holes and gasketing material forms a seal between the grids and the vessel walls. Overflow weirs control the bed depth and allows the sorbent to flow from grid to grid and to exit the vessels.

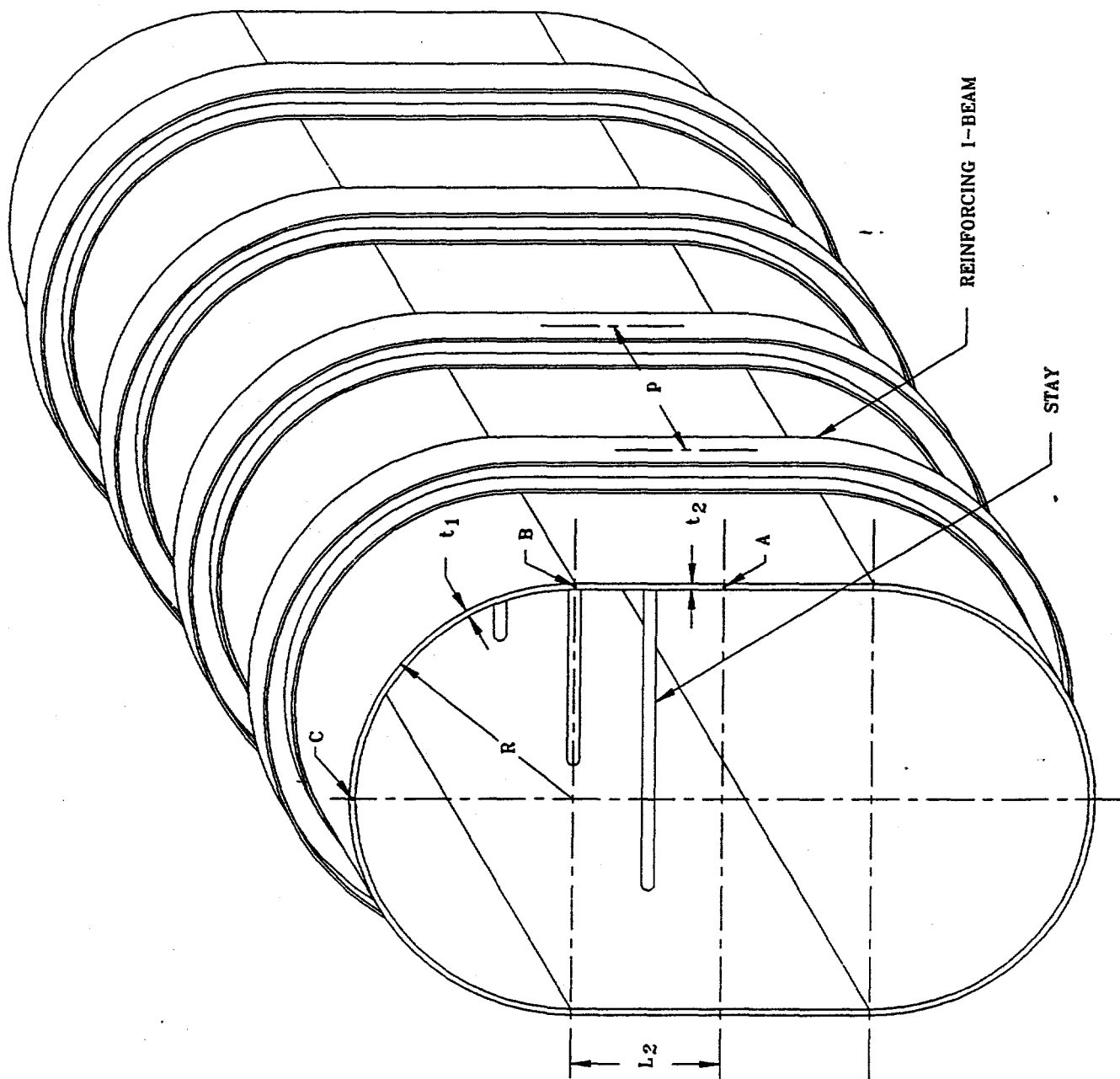


Figure 3-3. Horizontal Vessel of Obround Cross Section

Figure 3-4. Stress Distribution of Adsorber Vessel

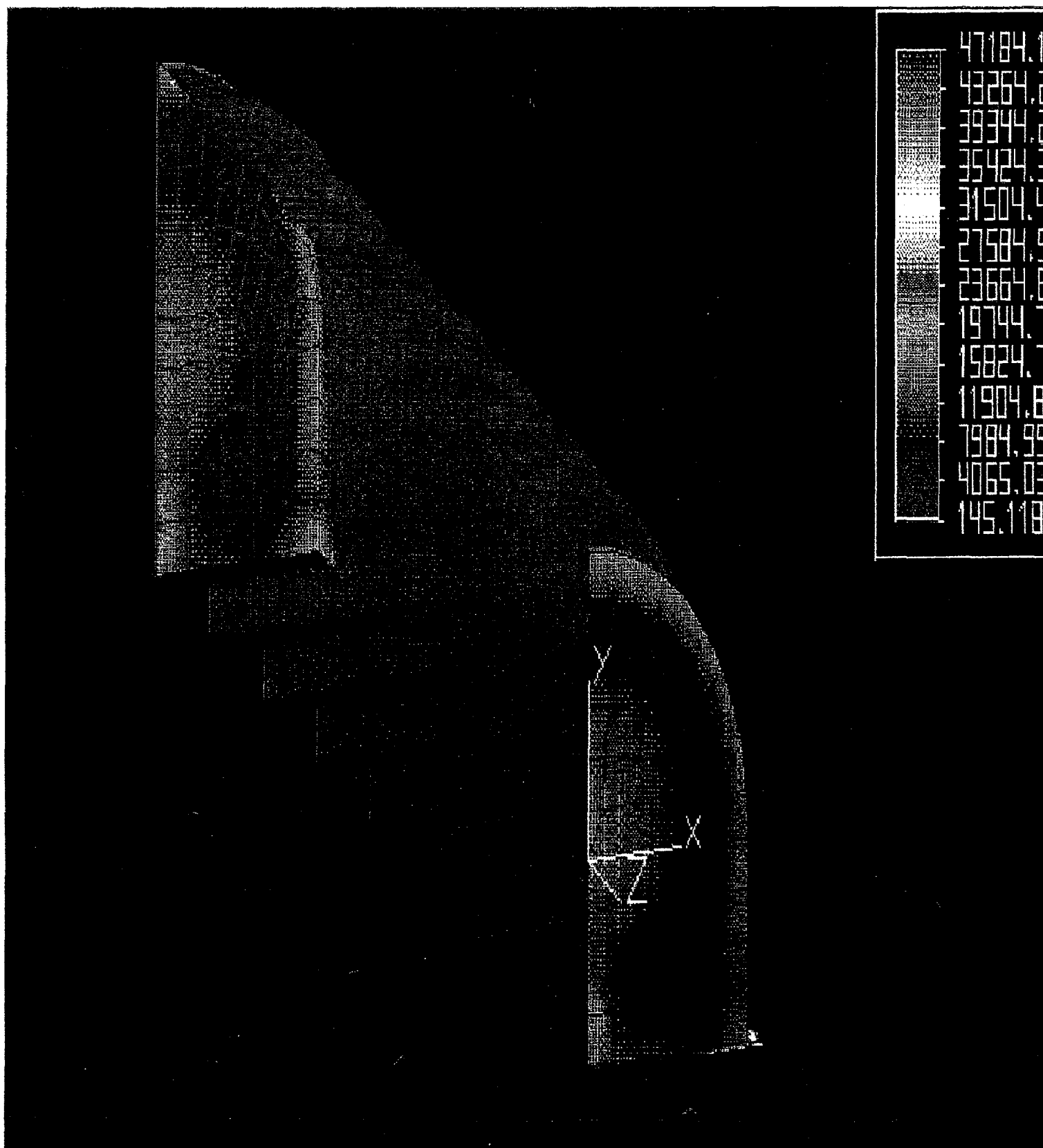
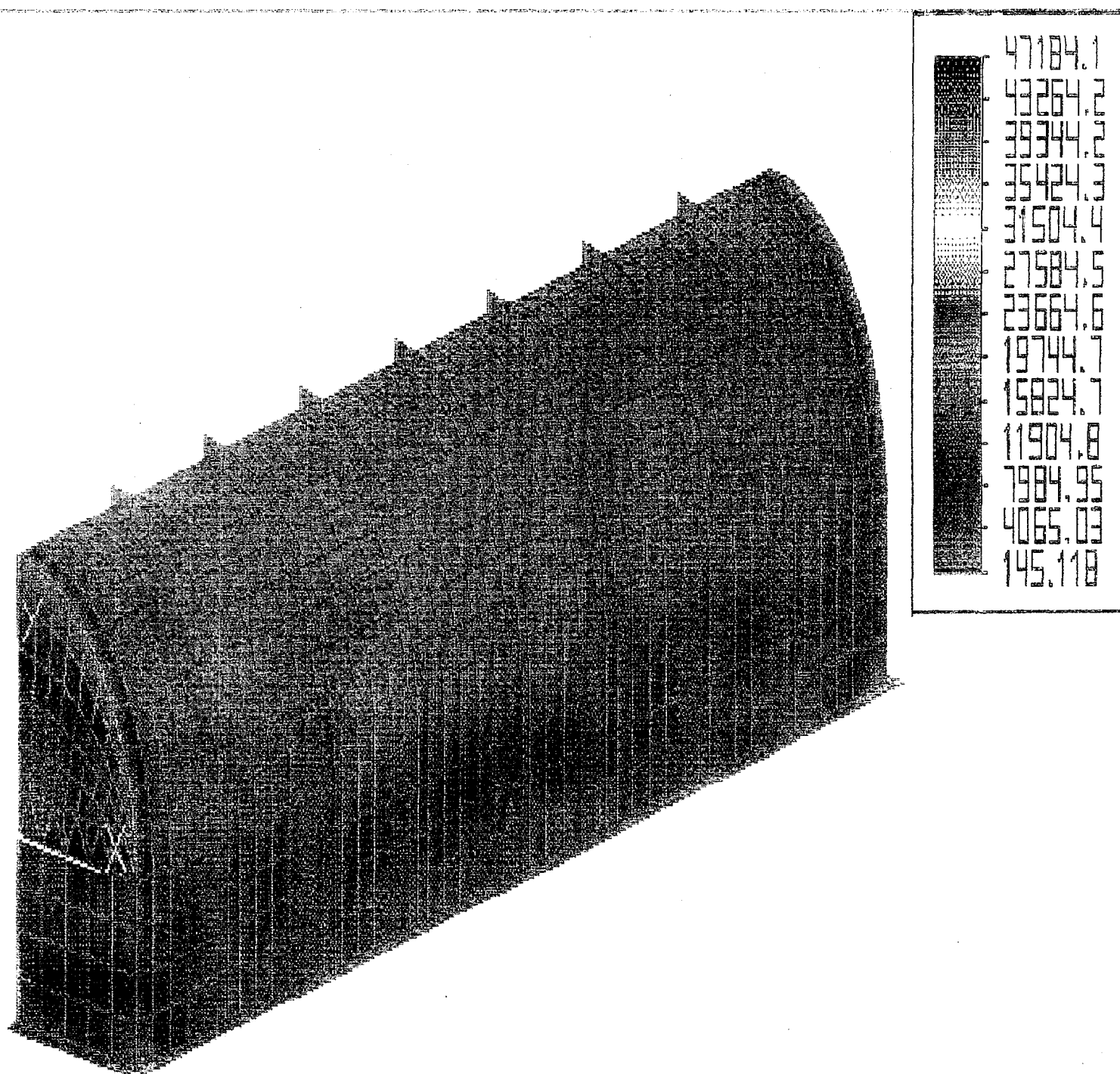


Figure 3-5. Stress Contour of Adsorber Vessel



While considerable reinforcement is required for a vessel of this cross section, the span of 12 feet reduces the structural requirements of the grid system as compared to the 34 foot cylindrical vessels. Heat transfer and the economic tradeoffs of the two vessel designs will be studied.

3.3.1.4 Vessel to Vessel Solids Transport

Dense phase transport systems are a key element in the design of the low profile plant. Sorbent must be transported between the four main operating units; adsorber, sorbent heater, regenerator, sorbent cooler, and back to the adsorber. Three different systems have been given serious consideration for use at the demonstration plant; J-valves, dense phase conveying and modified J-valves.

The POC successfully used J-valves to transport both hot sorbent (800°F to 1200°F) and cold sorbent (ambient to 300°F), and used a dense phase system to transport cold sorbent. These systems performed three functions; first, they transport sorbent with low attrition rates; second, provide pressure isolation between vessels operating at different pressure levels; and third, prevent the gas in a vessel from mixing with the gas in the other vessels.

J-valves operate in a fixed moving bed flow regime in the down flow leg using low volume counter current gas flow and operate in cocurrent dense or dilute phase flow in the up flow leg at high gas flow. POC experience has shown them to be effective in fluid transport, gas isolation, pressure isolation, and to have a low attrition rate. Two main difficulties were observed; the J valves exhibited a control hysteresis in that consistent control settings did not produce consistent results. In addition, proper design of the valves placed severe constraints on the relative lengths of the down flow and up flow legs. This in turn set vessel elevations and resulted in a high (200 feet plus) structure for the demonstration plant.

A dense phase system at the POC transported cold sorbent with repeatable control, provided a 70 feet lift, and exhibited low attrition levels. Special valves are required on the dense phase inlet to permit sorbent flow shutoff without producing high attrition. These valves have proven successful with cold sorbent, but different types of valves are required for hot sorbent. Two valve types have been proposed by different vendors, a water cooled valve with an inflatable seal ring, and an "everlasting" valve which employs a self-lapping rotating disk closure that is very expensive.

The main advantages of the dense phase system are good control and the proven ability to provide high sorbent lifts. This high lift means the vessels can now be located at low elevations. An additional advantage is the system is unaffected by pressure differentials between vessels.

Modified J-valves are the third method of sorbent transport. The bottom section of the J is squared off. This results in separation of the flow control from the lift gas and gives both repeatability and high lift (demonstrated at the POC.). When compared to a dense phase system, the drawbacks are a significant increase in utility consumptions when a high lift is required (not yet optimized) and increased down leg length required to operate with large pressure differentials and increased conveying velocity which translates into increased sorbent attrition.

The low profile design for the demonstration plant uses the dense phase systems for all sorbent transport.

3.3.1.5 Structural

The structural design for the low profile plant assumes generic site conditions. The facility consists of a NOXSO process tower, sulfur plant, MCC building, control and analyzer building, nitrogen storage tank, sulfur storage tank, and truck loading.

The NOXSO process tower contains several vessels, piping, platforms, duct work, and other equipment for cleaning the flue gas. It will be an open structural steel framed tower, approximately 48 feet high, with two stair towers. The steel frame is braced in both directions. The two major floors are covered with grating, and for safety there are hand rails all around each tower floor. The foundation of the process tower is designed on the assumption that a mat foundation will be the most suitable type for the future site. Soil bearing capacity assumes 3 kips per ft² and the bottom of the mat is to be 6'-0" below finished grade (0'-0").

Process tower floors and ground floor are designed for a minimum of 150 psf, live loads. The wind pressure is calculated assuming the design wind velocity is 80 mph, and exposure C with importance factor (I) as one.

The MCC building and control/analyzer building are enclosed structures. Roofs consist of hollow core pre-cast slab panels with load bearing masonry walls. Their foundation is part of the process tower mat. The ground floor is slab on grade.

The sulfur plant will provide most of the steam required for this project as well as treating the offgases from the NOXSO process tower. The plant and equipment are "skid mounted". The sulfur plant is supported on a mat foundation, principally because of the nature of the structure. The bottom of the mat edges are below the frost elevation.

The nitrogen storage tank, truck loading, and sulfur storage tank are supported on conventional spread footings.

3.3.1.6 Electrical Single Lines

The power requirement for the NOXSO process is 6.5 MW requiring a 1000 amp, 4160 volt, three phase feed to the NOXSO process switchgear. Motors 200 hp and larger shall be 4160 volt powered from the switchgear. These motors shall have WPII enclosures with heaters for outdoor use. Unit sub #1, fed from the 5kV switchgear, will provide 480 volt power to the low voltage motor control center and to the sulfur plant. The MCC will contain 480 volt motor starters and feeder breakers to distribution panels.

3.3.2 Sulfur Plant Technical Evaluation

The sulfur plant vendor selected has met the technical requirements as outlined in the request for quotation. The design is based on a range of H_2S to SO_2 ratios which have higher than "normal" concentrations of SO_2 . The excess SO_2 is hydrogenated to H_2S in the first processing step of the sulfur plant. Refer to **Figure 3-6**. Hydrogen for the reaction is produced by partial oxidation of methane in the presence of steam. Feed gas to the sulfur plant is cooled from $1000^{\circ}F$ to $600^{\circ}F$ and generates steam before combining with the reducing gas. The H_2S to SO_2 ratio is controlled at 2 moles H_2S to 1 mole SO_2 by varying the amount of reducing gas produced. The heat produced from the exothermic hydrogenation reaction is removed by a circulating hot oil loop eventually producing steam.

Two reaction stages are provided after the hydrogenation stage and the condensed sulfur drains to an above ground sulfur tank through a proprietary valve developed by the sulfur plant vendor. The condensing sulfur produces more steam for the steam system. The pressure and flow from the steam system can be varied during the final design to match the needs of the NOXSO process.

A booster blower is located between the two final reaction stages to provide the pressure drop required by the sulfur plant. The pressure at the outlet of this blower can be matched to the needs of the NOXSO process during final design.

Tail gas incineration is the last process step in the sulfur plant. H_2S in the tail gas is burned to produce SO_2 . The flue gas produced is then recycled to the NOXSO adsorber for SO_2 removal. (The NOXSO process acts as the sulfur plant tail gas clean up step). The percent sulfur conversion in the sulfur plant was balanced with the NOXSO process requirements for tail gas SO_2 removal to produce the most economical arrangement.

The sulfur plant will be controlled by the NOXSO distributed control system. The sulfur plant vendor will provide the control logic and the programming for their process.

The sulfur plant will be a skid mounted modularized unit. All instrument and control wiring will be terminated in a junction box at the module edge. Process and utility piping will be fabricated and flanged up to the battery unit.

3.4 Nitrogen Oxide Studies

No nitrogen oxide studies were conducted during this reporting period.

3.5 Process Studies

3.5.1 NOXSO Pilot Plant

The NOXSO pilot plant as originally constructed at Ohio Edison's Toronto Power Plant in Toronto, Ohio is shown schematically in **Figure 3-7**. Subsequent modifications (two stage adsorber and adsorber water sprays) are discussed in subsequent sections. A slip stream of flue gas (equivalent to 5 MW coal-fired power) is extracted from either Boiler #10 or #11 of Ohio Edison's Toronto Power Plant. The flue gas first flows through a 250 hp, F.D. fan, then is cooled to 160°C (320°F) or lower by spraying water into the flue gas ducts. The cooled flue gas then enters a 3.2 m (126") diameter fluid bed adsorber. The NOXSO sorbent, 1.23 mm diameter γ -alumina beads impregnated with 5.0 wt% Na, removes the SO₂ and NO_x simultaneously from the flue gas as it passes through the fluid bed adsorber. The cleaned flue gas mixes with the hot offgas from the sorbent heater, then passes through the baghouse to remove all particulate before the gas vents to the atmosphere through the power plant's chimney. A cyclone is installed after the adsorber to recycle fines to the bed (50% efficient on 50 micron diameter particles).

The spent sorbent in the adsorber overflows into the dense-phase conveying system where 377 kPa (40 psig) air lifts the sorbent 24.4 m (80 ft) high to the top of the sorbent heater, which is a 2.34 m (92") diameter, three-stage, fluid bed vessel. A natural gas-fired air heater supplies the hot air to heat the sorbent in the sorbent heater to 621°C (1150° F). During the sorbent heating process, all the adsorbed NO_x and a small portion of adsorbed SO₂ (1%) desorb from the sorbent. The hot sorbent heater offgas can either be vented to the atmosphere through the power plant's stack or mixed with the cleaned flue gas entering the baghouse. The hot sorbent in the bottom bed of the sorbent heater underflows into a J-valve. Either nitrogen or steam can be used to carry the sorbent through the J-valve into a 1.22-m (48") diameter moving-bed regenerator. Natural gas is the first regenerant to treat the hot sorbent in the regenerator at a temperature of 1130°F. The sulfate on the sorbent is reduced to SO₂, H₂S, and sulfide on the sorbent. Steam is then used to hydrolyze the sulfide to H₂S which occurs at a temperature of 1080°F. The offgas from the natural gas treater mixes with that from the steam treater before the combined stream enters the incinerator, in which all the sulfur species are converted to SO₂.

Regenerated sorbent flows into a second J-valve which conveys it to a 1.73 m (68") diameter, three-stage fluid bed cooler. A fan supplies ambient air to cool the sorbent. The heat of the regenerated sorbent is recovered by the cooling air which is then used as the combustion air in the air heater. The cool sorbent in the bottom bed of the cooler overflows into a 1.83m (72") diameter surge tank. A third J-valve is used to transport the sorbent from the surge tank to the adsorber.

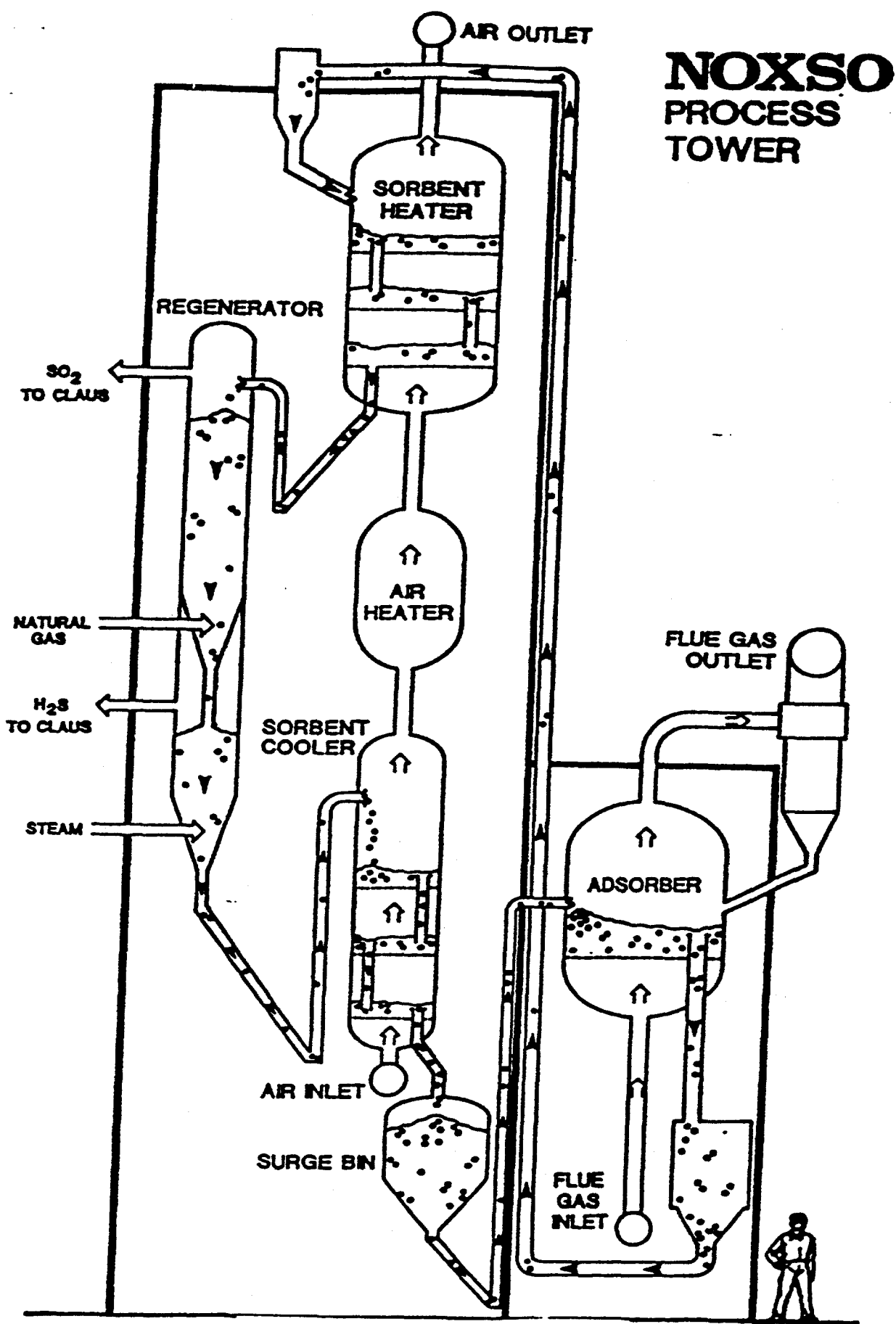


Figure 3-7. NOXSO Pilot Plant at Toronto, Ohio

The first two J-valves isolate the reducing environment (regenerator) of the NOXSO plant from its oxidizing environment (heater and cooler train). The steam (for normal operation) or nitrogen (for start-up) enters the two J-valves to carry the sorbent upward. Since the steam is introduced at the lowest point of the J-valve, which is also the highest pressure point between the two vessels, a steam barrier is created to prevent the mixing of the reducing gas with air or vice versa. The third J-valve is operated using air to lift the sorbent from the surge tank to the adsorber.

The NOXSO pilot plant differs from a commercial unit in two respects. First, since the POC plant uses only a slip stream of flue gas from the power plant, the amount of NO_x evolved from the sorbent heater is too small to affect the NO_x thermal equilibrium inside the boiler furnace. Therefore, NO_x is not recycled to the boiler as it would be in a commercial unit. However, the ability of a coal combustor to reduce excess NO_x introduced into the combustion chamber was proven in simulated NO_x recycle tests. The tests were carried out using the Department of Energy, Pittsburgh Energy Technology Center's tunnel furnace and 227 kg/hr (500 lb/hr) pulverized coal combustor.^{1,2} Additional NO_x recycle tests were conducted on a 227 kg/hr (500 lb/hr) small scale cyclone burner at the Babcock & Wilcox Research Center in Alliance, Ohio.³

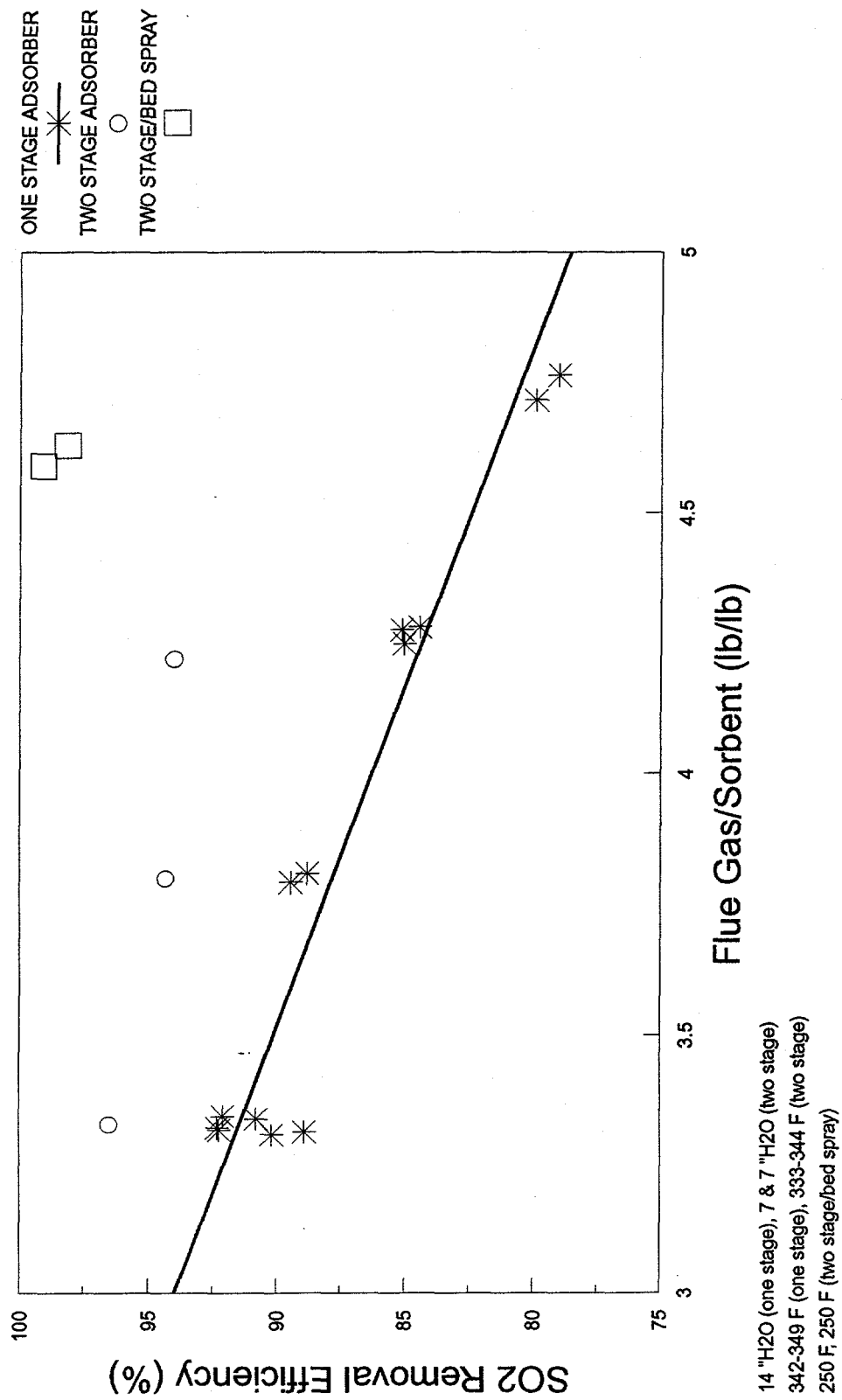
The second difference between the POC and a commercial NOXSO installation is the fate of the regeneration offgases. At the POC, the offgases are simply incinerated so that all sulfur species are converted to SO_2 and then vented to the plant stack. In a commercial unit, the regeneration offgases are sent to a sulfur recovery plant where the sulfur bearing gases are converted to elemental sulfur. Depending on market conditions and regional demand, either sulfuric acid or liquid SO_2 could be chosen as the end product rather than elemental sulfur. As the processes for making each of these end products are well established and commercially available, they were not tested at the POC facility.

3.5.1.1 SO_2/NO_x Removal Efficiency

In an effort to increase SO_2 and NO_x removal efficiencies, the pilot plant adsorber was modified to add a second bed and water sprays for in-bed cooling of the sorbent and flue gas. The performance of the two-stage adsorber and the in-bed water spray was evaluated in tests on actual flue gas at the Toronto Power Plant.

Baseline Testing

Baseline testing was performed to compare results obtained with a single stage adsorber. The two-stage testing was done at the same sorbent circulation rate, equivalent flue gas flows and equivalent flue gas temperature relative to single stage testing. Sorbent inventory for the single stage was split evenly between the two beds. Results are summarized in **Figure 3-8** for SO_2 removal and **Figure 3-9** for NO_x removal. These results met or exceeded expectations.



**Figure 3-8. NOXSO Pilot Test
One -Stage vs. Two -Stage Adsorber**

At conditions of 9000 scfm flue gas flow, 10,000 lb/hr sorbent circulation rate, 14 inch H₂O total bed pressure drop, and 340°F bed temperature, removal efficiencies for the single stage adsorber are 85% SO₂ and 71% NO_x. At these same conditions, the two-stage adsorber achieved 94% SO₂ removal and 89% NO_x removal. This improvement is attributed to improved gas- solids contacting in the two shallow beds versus the one deep bed since the second grid plate redistributes the gas before it enters the second bed. The improvement is also partially due to counter-current flow of gas and sorbent, since partially sulfated sorbent is exposed to the high concentrations of pollutants in the bottom bed and the exiting flue gas with the lower pollutant concentration is in contact with the fresh regenerated sorbent.

Parametric Testing - Variable Inventory

Due to the limited transport disengaging height of the second bed, only the bottom bed inventory was varied. The baseline testing was performed with a nominal 7 inch H₂O pressure drop per bed which corresponds to a settled bed height of 11 inches. The bottom bed was increased to a nominal 12 inch H₂O pressure drop or settled bed height of 18-3/4 inches.

Tests were conducted at conditions similar to two-stage baseline conditions for comparison. Table 3-1 summarizes test conditions and results for the two-stage baseline and variable inventory comparison. Higher SO₂ and NO_x removal efficiencies were obtained with the increased inventory on the bottom bed.

Table 3-1. NO _x /SO ₂ Performance with Variable Sorbent Inventory								
	Flue Gas (scfm)	Sorbent (lbs/hr)	Bottom Bed Ht (in)	Top Bed Ht (in)	Temp. Bottom Bed (°F)	Temp. Top Bed (°F)	SO ₂ Removal (%)	NO _x Removal (%)
Base	9121	9930	10.4	10.6	340	342	94.0	89.3
Variable Inventory	9194	9930	21.2	10.5	344	340	97.6	92.6
Base	8156	9924	11.0	10.7	342	345	93.7	90.6
Variable Inventory	8082	9938	18.8	12.8	344	342	98.9	96.1

Parametric Testing - Variable Temperature

Variable temperature testing was accomplished by spraying water into the sorbent beds. Spray nozzles were positioned above both beds and the water was sprayed into the bed. With this method, the flue gas is cooled while adsorption is taking place. Since it is known that the SO₃ is adsorbed very quickly, cooling below the acid dewpoint in the bed was not anticipated to be a problem as several inches above the grid plate there would be little SO₃ available to

combine with the water to form acid. The majority of testing with the water sprays was performed at temperatures below the acid dewpoint of the inlet flue gas. Inspections of the interior of the adsorber have revealed no acid corrosion as a result of spraying water into the sorbent beds.

Approximately 500 hours of water spray testing were performed. This was done to verify the expected increases in removal efficiencies and to determine if there was any increase in sorbent attrition with the water sprays. Table 3-2 compares two similar operating points without and with water sprays.

Table 3-2. NO _x /SO ₂ Performance with Variable Bed Temperature								
	Flue Gas (scfm)	Sorbent (lbs/hr)	Bottom Bed Ht (in)	Top Bed Ht (in)	Temp. Bottom Bed (°F)	Temp. Top Bed (°F)	SO ₂ Removal (%)	NO _x Removal (%)
without spray	9194	9930	21.2	10.5	344	340	97.6	92.6
with spray	9071	9937	18.7	11.1	241	265	99.0	95.7

Spiking Tests

Since the NO_x removal efficiencies with the two-stage adsorber were already very high at Toronto, the improvement with water sprays does not appear to be dramatic. The advantage of in-bed cooling with water sprays is best demonstrated when spiking to cyclone boiler concentrations. The Toronto flue gas SO₂ and NO_x concentrations were raised to simulate cyclone boiler flue gas by spiking the flue gas with SO₂ and NO_x from compressed gas cylinders. Over the duration of the POC project, 32 spiking tests were performed. Ten spiking tests were performed with the two-stage adsorber.

Table 3-3 compares three spiking tests at similar conditions. The adsorber configurations are single stage, two-stage, and two-stage with water sprays. The duration of the NO_x spiking for the single stage tests and two-stage test without water sprays was not sufficient to produce a steady state removal rate. Since the NO_x removal efficiency was dropping when these tests were concluded, the best that can be said is that the steady state removal efficiency would be less than the value measured at the end of the test. The duration of the NO_x spike for the two-stage test with water sprays was sufficient to reach a steady-state removal rate. The NO_x removal rate is strongly temperature dependent and cooling with water sprays produced a dramatic improvement, as shown in Table 3-3.

Table 3-3. NO _x /SO ₂ Performance and Adsorber Configuration								
	Flue Gas (scfm)	Sorbent (lb/hr)	Bottom Bed Ht (in)	Top Bed Ht (in)	Temp. Bottom Bed (°F)	Temp. Top Bed (°F)	SO ₂ Removal (%)	NO _x Removal (%)
Single Stage	9340	9900	33.6	-	328	-	88.6	63.4
Two-stage	9337	9882	21.6	10.5	317	326	97.8	48.2
Two-stage with water sprays	8963	9933	18.7	10.1	256	251	99.0	85.2

Data from three spike tests with a comparison to typical cyclone boiler design points is presented in Table 3-4. **Figures 3-10 and 3-11** are plots of SO₂ and NO_x removal efficiencies during spiking tests. Note that SO₂ removal increases with increasing NO_x concentration in the inlet flue gas. This substantiates one of the steps in the proposed reaction mechanism in which NO_x catalyses the oxidation of sulfite to sulfate, a stable compound.

Table 3-4. NO _x /SO ₂ Performance on Aged Sorbent/Cyclone Boiler Concentrations				
	Test #1	Test #2	Test #3	Typical Cyclone Boiler
Test Duration (hr)	3.5	1.7	2.7	-
Adsorber Temperature (°F)				
- top	308	251	252	250
- bottom	288	256	254	250
Adsorber Linear Velocity (ft/sec) @ Condition	2.4	2.3	2.0	2.8
Sorbent Residence Time (min)	52	45	46	52
Sorbent/Flue Gas Ratio (lb/lb)	0.24	0.24	0.20	0.24
NO Concentration (ppm)	917	1000	1167	971
SO ₂ Concentration (ppm)	2580	2365	2635	2650
Adsorber NO _x Removal (%)	87.5	85.2	86.0-90.7	87.9
Adsorber SO ₂ Removal (%)	97.4	99	99	98.7

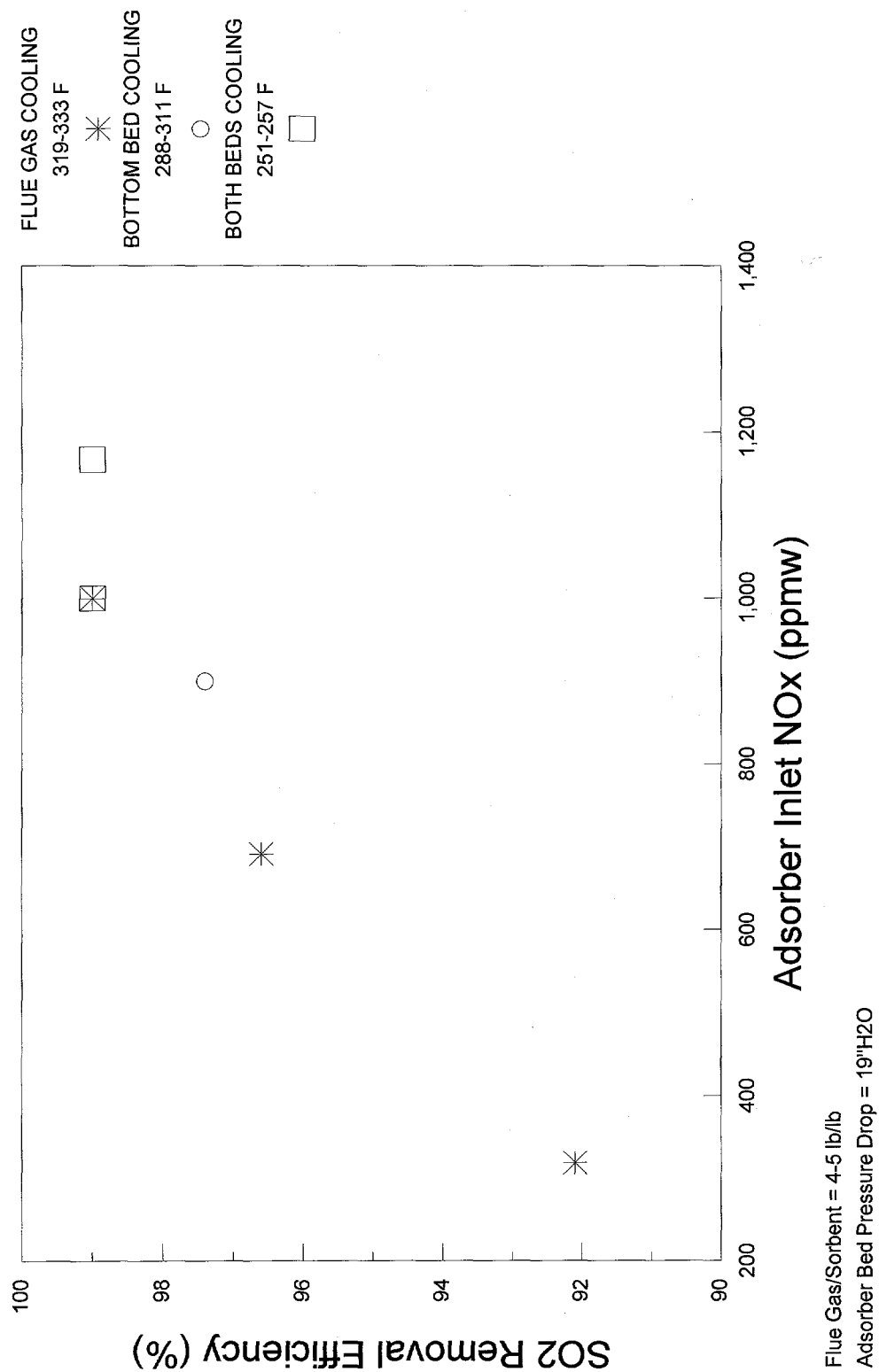


Figure 3-10. Flue Gas Spike Test: SO₂ Removal Efficiency as a Function of Inlet SO₂

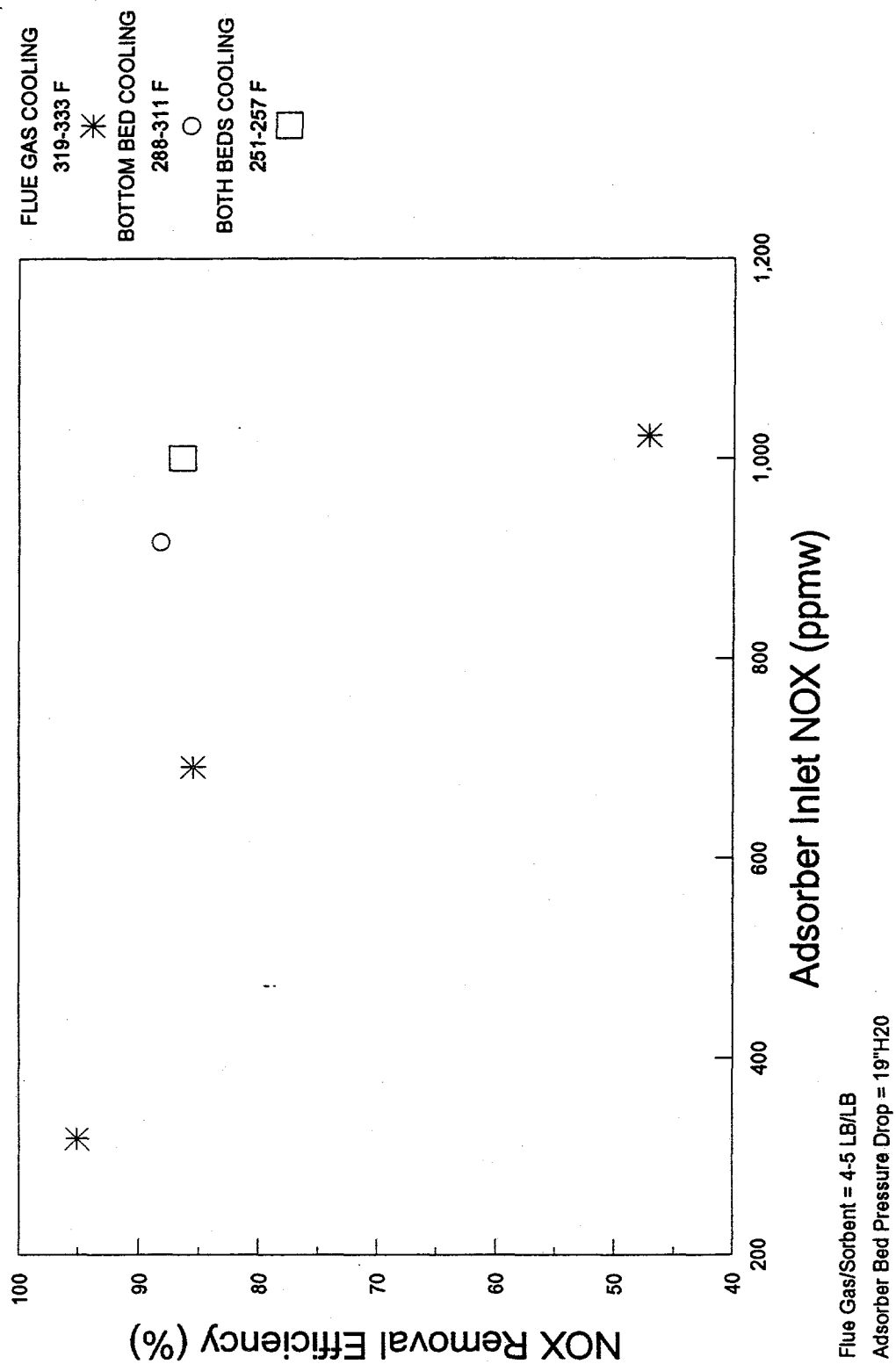


Figure 3-11. NOXSO Pilot Flue Gas Spike Test; NO_x Removal Efficiency as a Function of Inlet SO₂ Concentration

3.5.1.2 SO₂ Recycle Flue Gas System

On April 30, 1993, Ohio Edison closed the Toronto Power Plant thereby eliminating the source of flue gas for the NOXSO pilot plant. In order to continue testing, the pilot plant was modified to accomplish SO₂ recycle. In normal operation on flue gas, the sulfur exiting the sorbent regenerator in the form of SO₂ and H₂S is incinerated to SO₂ and exhausted to the power plant chimney. To prepare the plant for SO₂ recycle, the existing incinerator exhaust duct was extended a distance of approximately 30 feet and joined to the duct feeding the adsorber fan at the base of the north wall of the Toronto Power Plant. This junction was in the shape of a "T" with one leg of the "T" containing a damper and open to ambient air. In this way air could be added to control the temperature of the gas recycled to the adsorber fan.

One ton cylinders of liquid SO₂ were used to deliver the initial load of SO₂ to the system and to makeup SO₂ that was lost from the system in the adsorber offgas. The gas recycled to the adsorber was essentially air containing 1500-2500 ppm SO₂ with trace amounts of combustion products from the incinerator, i.e., CO₂ and water.

Tests on SO₂ recycle were conducted to obtain longer-term attrition data on the low density sorbent and to evaluate the affect of in-bed water spray on both the sorbent and the materials of construction in the adsorber. The value of adsorber performance data in these tests is limited since: 1) the SO₂ recycle does not contain NO_x, 2) the adsorption chemistry is different because of the differences in gas composition between SO₂ recycle and flue gas, and 3) the low density sorbent used in these tests was relatively "fresh" and tests have shown the performance of the "fresh" sorbent to be superior to aged sorbent. Nevertheless, the data show that the sorbent adsorbs SO₂ from a gas stream containing air, 1500-2500 ppm SO₂, and trace amounts of CO₂ and water.

3.5.1.3 Sorbent Attrition

Sorbent attrition was measured in two ways: by weighing material collected in the baghouse and by weighing the amount of sorbent makeup to the system to maintain a constant inventory. Experience has shown that the sorbent makeup rate is the most accurate measure of sorbent attrition. The baghouse collection data is of limited use since 1) the material collected in the baghouse is a mixture of flyash and sorbent (in tests on flue gas) and 2) the baghouse was not leak-tight, i.e., it was not possible to completely seal dampers in the baghouse bypass line or the pressure relief valve in the baghouse inlet line. For these reasons, the sorbent makeup rate was always higher than the baghouse collection rate.

Two sorbents were tested at the NOXSO pilot plant. The sorbents were chemically similar; the only difference between the two was the bulk density. The first sorbent tested had a bulk density of 42 lbs/cu.ft. The second sorbent tested was made via a different manufacturing process and had a bulk density of 34 lbs/cu.ft. Because of the difference in the manufacturing process, the cost to produce the lower density sorbent is less than the cost of the higher density sorbent.

During single stage adsorber testing, the sorbent makeup rate for the higher density sorbent measured over a period of 3232 operating hours on flue gas at the NOXSO pilot plant was 3.0 PPH.

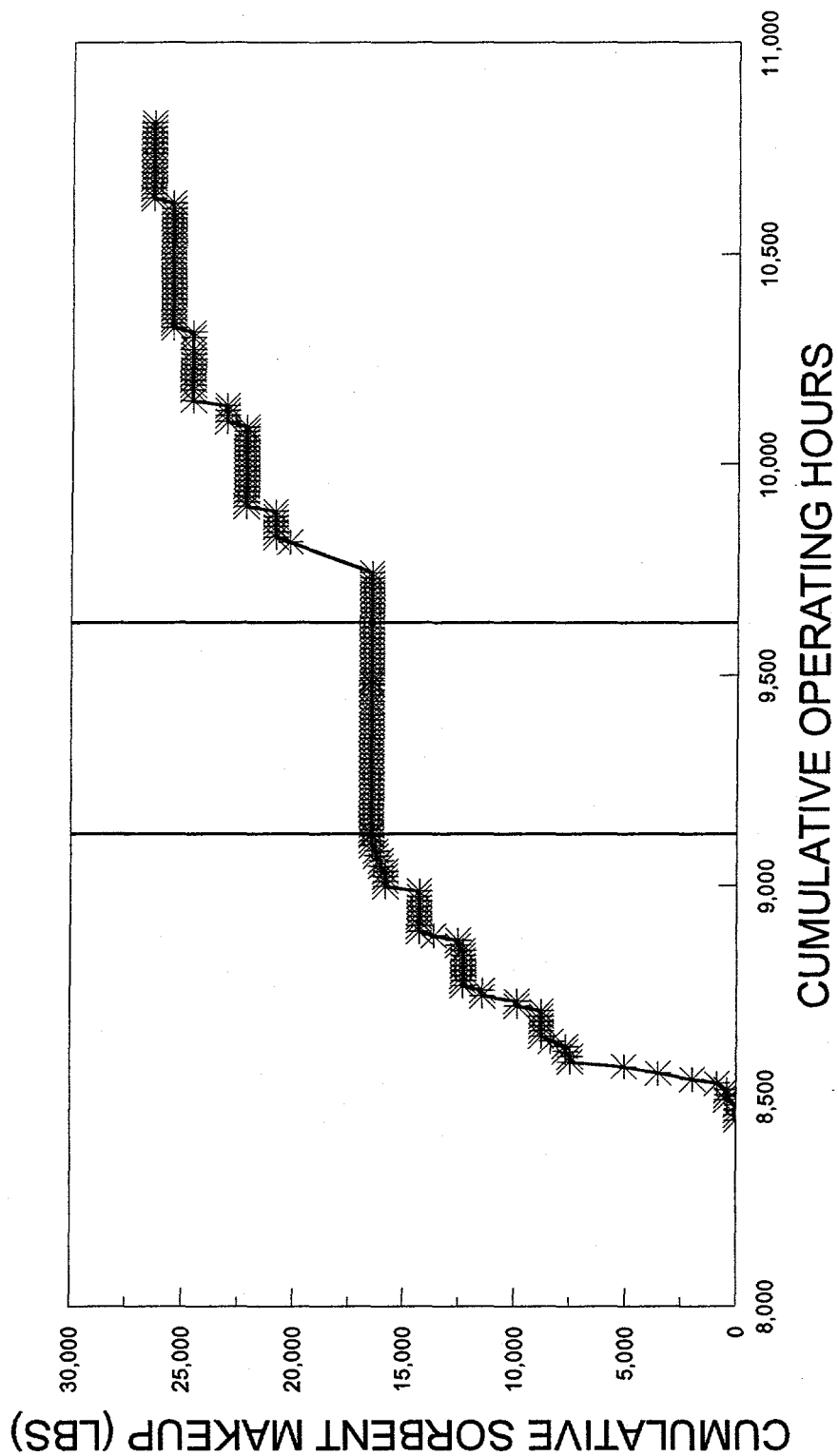
Figure 3-12 is a plot of cumulative operating hours versus cumulative sorbent makeup rate for the test on the lower density sorbent. The slope of this graph is the sorbent makeup rate, and as can be seen from the plot, it varies considerably during the test. One reason for this is that prior to 9123 operating hours (the period to the left of the heavy dotted line in **Figure 3-12**) tests were run on the lower density sorbent in the two-stage fluid bed adsorber. After 9123 operating hours, the tests were run on the single-stage adsorber. As seen in **Figure 3-12**, the sorbent makeup rate dropped considerably between the two-stage and the one-stage tests.

The sorbent makeup rate for the two-stage tests is higher because the two-stage adsorber had insufficient transport disengaging height (TDH) above the top bed, since the adsorber at the pilot plant was originally designed for only one stage. The insufficient TDH caused excessive carryover of sorbent from the top bed of the adsorber. Therefore, it is not possible to separate what is attrition and what is carryover in the two-stage tests.

There are two dotted lines shown in **Figure 3-12**. The line at 9123 hours marks the beginning of the one stage tests; the line at 9624 hours marks the beginning of one-stage tests with in-bed water spray. The attrition rate is determined by the sorbent makeup rate (the slope of the line in **Figure 3-12**), the starting and ending system inventories, and adjustments to inventory due to sample taking, accidental sorbent discharge, etc. The attrition rates are shown in Table 3-5. The relatively low attrition rate measured for the one-stage test before in-bed water spray may be due to the fact that this test was run on air before SO₂ recycle began. On air, the superficial gas velocity in the adsorber was about 15% lower than that in the SO₂ recycle tests because of differences in gas temperature. It should also be noted that at 9743 hours, shortly after in-bed water spray was begun, an L-valve was installed between the sorbent heater and regenerator. The L-valve is a potential source of sorbent attrition, but it is impossible to separate the effect of the L-valve from the effect of the in-bed water spray.

Table 3-5. Sorbent Attrition		
Sorbent	Test	Attrition Rate (PPH)
High Density	Single-stage adsorber	3.0
Low Density (9123-9611)*	Single-stage adsorber air through adsorber	3.3
Low Density (9624-10,634)	Single-stage adsorber SO ₂ recycle through adsorber in-bed water spray L-valve installed	5.5

*Operating hours, refer to Figure 3-12.



**Figure 3-12. NOXSO Pilot Test
Sorbent Makeup Rate
Low Density Sorbent (34 lbs/cu.ft)**

3.5.1.4 In-Bed Water Spray

Since NO_x removal efficiency increases with decreasing adsorber bed temperature (as shown previously), in certain applications of the process it may be desirable to cool the adsorber bed. Two methods of cooling were tried at the pilot plant: flue gas cooling via water spray into the ductwork approximately 90 feet upstream of the adsorber and cooling of the flue gas and bed simultaneously by spraying water directly into the bed from a nozzle placed above the bed. The latter option is preferable since: 1) the acid dewpoint in the bed is lower than in the flue gas, therefore the adsorber can be made of less expensive materials of construction, and 2) the flue gas upstream of the adsorber can be maintained above the acid dewpoint so that specialty acid-resistant materials are not required in the flue gas ductwork.

The in-bed water spray nozzle was positioned about 48 inches above the expanded fluidized bed in the center of the circular adsorber. The spray nozzle delivers a coarse spray that results in approximately 94% of the water reaching the sorbent bed with the remainder being carried away with the flue gas. The height of the spray nozzle above the bed was adjusted to give a spray whose perimeter was 6 inches from the vessel wall. The nozzle employed a swirl action which made it self-cleaning.

The adsorber at the pilot plant was made of carbon steel with acid-resistant linings on the bottom shell and in the flue gas ductwork after the in-duct water spray. The adsorber internals consisted of a Hastalloy bottom grid, a 304 stainless steel top grid, and a carbon steel overflow and downcomer. The adsorber was inspected at the end of the pilot test after 7000 hours of operation on flue gas including 1100 hours of operation using the in-bed water spray. The inspection uncovered no evidence of acid corrosion in the adsorber. This result means that the adsorber can be constructed of carbon steel with an appropriate corrosion allowance. The carbon steel corrosion coupons at the adsorber inlet yielded a corrosion rate of 3.8 mils/yr (1 mil = 0.001 inch). With a 30 year lifetime, a carbon steel duct would loose 0.114 inches of thickness. Carbon steel coupons in the adsorber corroded at 2.1 mils/yr.

3.5.1.5 Corrosion and Materials

Refractory samples of bricks and mortar were placed into the regenerator in February 1993. These samples were placed into stainless steel trays attached to the upper and lower manways of the regenerator. The upper regenerator samples were in the gas spaces of the regenerator while the lower regenerator samples were in the sorbent. The samples were exposed in the regenerator for a total of 3337 total hours (regeneration and hot inert) and 2433 total regeneration hours. The sample of HB mortar from the upper regenerator showed visible cracks around the entire cylindrical sample. The K14 mortar at this location looked excellent (as did the HB mortar in the lower regenerator). All of the brick samples looked excellent (with only minor weight changes and no visible erosion, cracking, or chemical attack of the samples). Table 3-6 summarizes the sample weight changes measured on the samples.

Several spray coated metallic samples were also tested in the upper regenerator. These samples were spray coated with C-276, Hastalloy C, METCO HOV alloy, and 45 CT alloy. All of the spray coatings peeled off the coupons during the exposure period. The C-276, Hastalloy C, METCO HOV alloy, and 45 CT alloy all showed very substantial attack of the metal spray coat.

Corrosion coupon samples were placed into the adsorber to measure the corrosion rates in the fluid bed directly below the water spray. This water spray was added to cool the lower sorbent bed of the adsorber by spraying water directly into it. The corrosion coupon rack was exposed in the vessel between May and August 1993. This rack contained the same types of coupons used to measure the corrosion rates entering and within the adsorber. (These results are summarized in the NOXSO POC Final Report). The coupons were exposed to a total of 1524 hours in the sorbent bed, with the majority of the operating time using the in-bed water spray. The results of this test are listed in Table 3-7. The units in this table are mils of loss/year of operation. With the exception of the teflon samples, the results are similar to the earlier measurements of corrosion at the adsorber inlet. The 446 stainless showed the highest rate of metal loss (1.3 mils/yr) followed by the C1010 carbon steel (0.8 mils/yr). All of the other specialty metals showed no significant corrosion (less than or equal to 0.1 mils/yr). The teflon samples showed substantial loss of material (1.9 and 2.7 mils/yr). This is due to the softness of the teflon and erosion of the teflon surface by contact with the fluid bed.

Vessel inspections were performed in August by NOXSO, W.R. Grace, and a W.R. Grace consultant (Dr. Van Sciver). The inspections showed the vessels to be only slightly changed from the inspection in February 1993. No significant metal loss was observed in any of the vessels. Several small pits were observed in the regenerator, but these are not considered significant at this time. The unalloyed 304 SS sorbent inlet deflector at the top of the regenerator shows substantial corrosion (as expected). The adsorber vessel was in excellent condition, but the Martek coating used to protect the bottom shell of the vessel has failed in several areas. Since this vessel is carbon steel, this coating should be replaced if the vessel is used for additional flue gas service. A thin layer of sorbent/ash had coated the bottom of the grid for the upper adsorber bed. This apparently formed from water/sorbent/ash but had not plugged the grid holes for the upper grid.

**Table 3-6. Ceramic Samples Placed in Regenerator for Testing 2433 Flue Gas Hours Total,
3337 Total Hours Operation (Flue Gas + Inert)**

Coupon	Location	% Wt loss
THERMAX XES, #1	REG UPPER	-1.2
THERMAX XES, #2	REG LOWER	0.0
VISIL, #1	REG UPPER	0.1
VISIL, #2	REG LOWER	0.0
KOCH AP300, #1	REG UPPER	0.5
KOCH AP300, #2	REG LOWER	0.0
KALA, #1	REG UPPER	0.0
KALA, #2	REG LOWER	0.0
K14 MORTAR, #1	REG UPPER	4.8
K14 MORTAR, #2	REG UPPER	4.7
K14 MORTAR, #3	REG LOWER	2.0
K14 MORTAR, #4	REG LOWER	2.5
HB MORTAR, #7	REG UPPER	7.2
HB MORTAR, #8	REG UPPER	9.2
HB MORTAR, #9	REG LOWER	4.8
HB MORTAR, #10	REG LOWER	4.5
Metal Sprayed Coupons		
SPRAYED, HASTALLOY C, "A"	REG UPPER	12.2
SPRAYED, C-276, "B"	REG UPPER	11.2
SPRAYED, 45-CT "K"	REG UPPER	9.5
SPRAYED, METCO HOV, "F"	REG UPPER	27.7
SPRAYED, METCO HOV, "FA"	REG UPPER	4.2

Table 3-7. Adsorber Inlet Coupon Rack for In Bed Corrosion of Samples 1524 Hours Total Exposure		
Coupon	Location	Loss mils/yr
C4, 3	AI 1	-0.1
C4,4	AI 13	0.1
C276, HT G825, A0167	AI 2	0.0
C276, HT G825, A0168	AI 14	0.0
316, HT 731, A0229	AI 3	0.0
316, HT 731, A0230	AI 15	0.0
INCO, 625, 3	AI 4	0.0
ICNO, 625, 4	AI 16	0.0
20 CB3, HTG774, A0089	AI 5	0.1
20CB3, HTG774, A0090	AI 17	0.1
JESSOP C276, 3	AI 6	0.1
JESSOP C276, 4	AI 18	0.1
C22, HT F120, A0055	AI 7	0.1
C22, HT F120, A0056	AI 19	0.0
C1010, HT F101, A0015	AI 8	1.1
C1010, HT F101, A0016	AI 20	0.4
304, 828	AI 9	0.1
304, 829	AI 21	-0.1
INCO C276, 3	AI 10	0.0
INCO C276, 4	AI 22	0.0
446, 3	AI 11	1.2
446, 4	AI 23	1.4
TEFLON, 3	AI 12	2.7
TEFLON, 4	AI 24	1.9

3.5.2 Demonstration Plant

Work was performed this quarter on several major aspects of the demonstration plant design. These include development of an adsorber model, development of a computer process simulator, technical and market study of sulfur by-product options, HCl removal by the process, and correlating L-valve experimental data with design equations. Each of these topics is discussed below.

3.5.2.1 Adsorber Model

A mathematical model for the fluid-bed adsorber was developed from the earlier NOXSO test results. The sorbent capacity for the NO_x and SO_2 sorption was determined from the laboratory using a two inch fixed-bed reactor. The rate constants of the NO_x and SO_2 sorption used in the model were derived from the NOXSO PDU tests. These constants are global rate constants, which lump the hydrodynamics of gas-solid contact and the reaction kinetic mechanism into one variable. This model had been tested with the LCTU adsorber and POC adsorber data. It was found that the PDU adsorber model accurately predicted the LCTU and POC adsorption results when the change of sorbent surface area and the difference of sorbent properties were incorporated into the model. However, the use of global rate constants oversimplifies the complicated adsorption process occurring in a fluid bed. Optimizing the adsorber design is unlikely using this model.

During the POC test, a statistical experimental program was planned and executed to develop an empirical adsorber model. The intent was to model SO_2 and NO_x removal efficiencies as a function of the design variables, so that optimum conditions and process tradeoffs could be evaluated. The experimental design was a thirty run central composite response surface design. Regression analysis was used to develop empirical prediction models for NO_x and SO_2 removals. The equations effectively represent the thirty run data base: 95% of the variability in NO_x removal, and 97% of that in SO_2 removal. Since the model is empirical, it is useful for interpolating the data base but is not suitable for extrapolating the results. It is unwise to apply the model to design a plant which is outside the range of the thirty run data base.

In order to optimize the NOXSO process, a more sophisticated reactor model is needed for reactor design. The reactor model should be able to simulate the real reactor of different sizes under different operating conditions.

Model Description

The NOXSO process is a heterogenous non-catalytic gas-solid reaction. The gas species diffuse into the sorbent then adsorb and react with the solid component on the sorbent substrate. In the adsorption step, the product species stay on the sorbent surface. In the regeneration, the product species desorb and diffuse out of the sorbent. For a fluid-bed adsorber, the reaction can be considered isothermal. Since the pollutant concentration of the flue gas is in the range of a couple thousand parts per million, the change of gas volume is insignificant. Therefore, it is

unnecessary to include the temperature and pressure dependence into the adsorption model. But the situation of regeneration is totally different. To demand a high methane usage, the change of gas volume is significant. The temperature and pressure profile across a moving-bed reactor can not be uniform. The model has to include the temperature and pressure.

The strategy of the model development is to develop a computer program to simulate the non-catalytic gas-solid reaction then correlate the program results with the test results to determine the various rate constants. During the program development, literature results will be used to check the computer outcome of the model's limiting cases to confirm the programming. Because reactions in the NOXSO process are very complicated, it is unlikely to find any similar problems which have been solved in the literature.

The current status of the model development is to develop the program core and then compare the computer results with the laboratory fixed-bed adsorption tests. The governing equations for the one dimensional fixed-bed reactor are given as follows:

Gas Phase

$$\begin{aligned}\frac{d}{dw} (F_g C_i) &= -6 \frac{k_{ci}}{d_p \rho_s} (C_i - C_i^s) \\ \frac{d}{dw} (C_{pg} \rho_g F_g T) &= -\frac{6 h_g}{d_p \rho_s} (T - T_s) \\ \frac{dp}{dw} &= -\frac{1-\epsilon}{\epsilon^3 \phi_s d_p A^2 \rho_B g_c} \left\{ [150 \frac{1-\epsilon}{\phi_s d_p}] \mu F_g + \left(\frac{1.75}{A} \right) \rho_{gz} F_g^2 \right\}\end{aligned}$$

Solid Phase

$$\begin{aligned}\frac{1}{\xi^2} \frac{d}{d\xi} \left(\xi^2 \frac{dC_i^s}{d\xi} \right) &= \frac{R^2 S A \rho_s}{D_{ei}} \gamma_i \\ \frac{dT_s}{dt} &= -\frac{6}{d_p \rho_s} \left\{ h_g (T_s - T) + \sum (\Delta H_i) \kappa_{ci} (C_i - C_i^s) \right\} / C_{ps} \\ \frac{dC_{sj}}{dt} &= f(\kappa_k, C_i^s, C_{sj})\end{aligned}$$

Boundary and Initial Conditions

$$\begin{aligned}
 &\text{For } t=0, \text{ all } W, T=T_0 \text{ and } T_s=T_{s0} \\
 &\text{For } W=0, \text{ all } t, C_i=C_{i0}, C_{sj}=C_{sj0} \text{ and } F_g=F_{g0} \\
 &\text{For } t=0, \text{ all } W, \xi, C_{sj}=C_{sj0} \\
 &\text{For } \xi=0 \text{ all } t, W \quad \frac{dC_i^s}{d\xi} = 0 \\
 &\text{For } \xi=1 \text{ all } t, W \quad \frac{2D_{ei}}{d_p} \frac{dC_i^s}{d\xi} = k_{ci}(C_i - C_i^s)
 \end{aligned}$$

The program is written in Microsoft FORTRAN, and tested on an IBM compatible 486 PC.

Nomenclature

A	cross-sectional area of fixed-bed reactor, cm ²
C _i	gas concentration in the gas phase, gmole/cm ³
C _i ^s	gas concentration inside the sorbent pellet, gmole/cm ³
C _{pg}	gas heat capacity, cal/gmole.°C
C _{sj}	solid concentration, gmole/gsorbent
D _{ei}	internal effective diffusivity, cm ² /sec
d _p	sorbent diameter, cm
F _g	gas volumetric flow rate, cm ³ /sec
g _c	gravitational conversion factor
h _g	external heat-transfer coefficient, cal/cm ² .sec.°C
k _{ci}	external mass-transfer coefficient, cm/sec
P	pressure, atm
SA	sorbent surface area, cm ² /gsorbent
T	gas temperature, °C
T _s	sorbent temperature, °C
t	time, sec
W	sorbent weight, g
γ _i	rate of reaction, mole/sec·cm ²
ΔH _i	heat of reaction, cal/gmole
ε	fixed-bed voidage
κ _k	reaction rate constant of kth reaction, the unit of rate constant is dependent on the rate form.
μ	gas viscosity, g/cm·sec
ξ	dimensionless particle radius
ρ _B	sorbent bulk density, g/cm ³
ρ _g	gas molar density, gmole/cm ³

ρ_{gz} gas mass density, g/cm³
 ρ_s sorbent particle density, g/cm³
 ϕ_s sorbent particle sphericity

subscript i stands for the ith gas species
subscript j stands for the jth solid species

3.5.2.2 Process Simulation

To effectively and rapidly model the NOXSO process, a computer simulation of the NOXSO process was developed. The process simulation model is a modularized Quick Basic program consisting of over 3000 lines of code and 20 subroutines. The heart of the program is the fluid bed adsorber subroutine. The fluid bed adsorber design equation used in the routine was developed to model the Life Cycle Tests Unit and 3/4 MW test data. The equation was later adapted to model the Proof-of-Concept test data. A new, more advanced adsorber model is currently being developed as discussed previously. The moving bed regenerator routine is an empirical model based on Proof-of-Concept data.

Given information for a specific coal fired power plant, the process simulation will model a NOXSO combined SO₂/NO_x removal system integrated with the power plant. Site specific information required includes the following: plant gross generating capacity, coal feed rate, coal analysis, flue gas composition, flue gas temperature, and combustion air heater in leakage. Other inputs to the program customize and optimize the basic NOXSO process for a given power plant. These inputs can include the following: desired SO₂ removal efficiency, number of stages in the sorbent heater and sorbent cooler, adsorption temperature, and sorbent sulfur loading.

The process simulation output contains three main sections: equipment, stream data, and process economics. The equipment section includes the power requirements for all rotating equipment. This includes the adsorber fans, sorbent heater/cooler train fans, sulfur recovery unit fans, and the air compressors. All major vessels are sized, and where applicable fluid bed depths in the vessels are also shown. The vessels sized include the following: adsorber, sorbent heater, sorbent cooler, methane treater, steam treater, and surge bin.

The stream data section contains the results of the calculated material, pressure and energy balances performed on a process flow diagram containing over 120 streams. Each stream contains the mass flow rates of sorbent and total gas, with the gas mass flow divided into 13 components; the temperature and pressure, the gas heat capacity, and the gas average molecular weight. The economics output contains the results of an analysis using inputted consumables costs. Total operating and maintenance costs are given. Revenue from excess SO₂ allowances and sulfur by-product recovery is shown and the bottom line levelized costs are presented.

3.5.2.3 Sulfur By-product Study

A technical and economic study of sulfur by-product options is ongoing. In addition to in-house work, a consultant familiar with the sulfur industry has been retained. The three sulfur by-products under review are: 1) molten sulfur, 2) sulfuric acid, and 3) liquid SO₂. This section will discuss the preliminary findings of this sulfur by-product study.

From a by-product value consideration, liquid SO₂ is favored. For every ton of sulfur produced, two tons of liquid SO₂ and 3.1 tons of 98.5% sulfuric acid can be produced. According to the Chemical Marketing Reporter for the week ending 6/21/93, the spot f.o.b./ton prices were as follows: sulfur \$44-\$53, liquid SO₂ \$230, and sulfuric acid \$52-\$75. However, the net prices received for these sulfur by-products is expected to be less than the spot prices due to shipping costs, transfer fees, and long term contracts to bulk users. Assuming a reasonable net price for the sulfur by-products, Table 3-8 compares the potential sulfur by-product revenue.

Table 3-8. Sulfur By-product Revenue			
	Production Rate*	Net Value	Net Revenue
Liquid Sulfur	7,598	\$25/ton	\$179,950
Sulfuric Acid	23,555	\$40/ton	\$942,200
Liquid SO ₂	15,197	\$120/ton	\$1,823,640

*Based on a 100 MW plant burning 3% S coal, assuming a 70% capacity factor and 95% SO₂ removal efficiency.

Of concern to the involuntary producer, or any producer of a marketable commodity, is the market volatility and size. Currently the market for liquid SO₂ is stable, however the market for sulfur and sulfuric acid is very cyclic. The sulfur and sulfuric acid market is currently in decline after reaching a peak in 1985, with market forecasts appearing gloomy. According to one industry publication, the depressed market is due in large part to the sulfur and sulfuric acid industry changing from voluntary to involuntary producers. Involuntary producers generate sulfur by-products as a waste stream as NOXSO would do. The metal smelter industry is another example. They are now required by environmental regulations to recover large quantities of sulfur previously released to the atmosphere as sulfur dioxide.⁴ Eventually, most voluntary producers of sulfur by-products will be forced out of business by involuntary producers.

Sulfur and sulfuric acid are major commodity chemicals and the market could easily absorb the NOXSO demonstration units production. However, the North American SO₂ market is relatively small at 600,000 TPY. (A NOXSO demonstration unit in the 100 MW size range would supply 2.5% of the total North American market.) The market is divided into three categories: 1) liquid merchant, 2) liquid captive, and 3) gas captive. The liquid merchant category, primarily a U.S. market, produces approximately half of the total market production

of SO₂. Within the liquid merchant category, approximately 100,000 TPY of liquid SO₂ is produced by voluntary producers who primarily burn sulfur. The balance is supplied by involuntary producers who generate liquid SO₂ as a by-product waste stream. The liquid captive category, users who have installed equipment at the point of use to produce liquid SO₂, accounts for 75-85,000 TPY of the total North American market. The last category, the gas captive market, accounts for 200-210,000 TPY of the total market. Producers there are primarily Canadians who produce gaseous SO₂ at the point of use.

From a market economic stand point, there is not a clear choice among the three sulfur by-product options. The sulfur and sulfuric acid market is sufficiently large to absorb the NOXSO demonstration unit's production but with falling prices the sulfur or sulfuric acid may be of little or no value. Conversely, the SO₂ market is small but offers a stable, relatively high priced market. In the SO₂ market, NOXSO would be competing with other involuntary users in the liquid merchant category, and if the price fell to \$50-\$60/ton could replace voluntary producers within this category.⁵ Perhaps the overriding factor determining the sulfur by-product choice will be the host sites location and/or needs. Shipping costs can become excessive in relation to market price, especially for sulfur and sulfuric acid. If the host site is an end user or near an end user of one of the sulfur by-products, this sulfur by-product could be the clear choice.

The technical status of the sulfur by-product study is at varying levels of completion depending on the by-product. The current NOXSO design includes a sulfur plant. To this end, a large amount of engineering work has been performed to determine the feasibility of producing molten sulfur and NOXSO is confident of the industry's ability to produce molten sulfur from the regenerator offgas. Most of the early sulfur plant design work was performed by The Ralph M. Parsons Company for a sulfur plant at the now canceled Niles location. Advanced Petrogas Systems (APS) more recently submitted a quote with the same process parameters. APS design and fabricate most of the large vessels and equipment in-house. Additionally, the APS plant would be skid mounted and thus reduce more costly and less efficient field labor. Parsons, on the other hand, subcontracts the design and construction of most of the sulfur plant and relies on field construction to erect the sulfur plant.

Chemetics International Company was contacted by NOXSO concerning sulfuric acid and liquid SO₂ production from the regenerator offgas. They provided a preliminary design for a field constructed 150 TPD, single adsorption contact sulfuric acid plant. According to Chemetics, this plant size is at the bottom end of their design range. Thus, economies of scale would work in favor of a larger, say 500 MW, NOXSO unit. Currently, a sulfuric acid bid specification package has been prepared and is in the process of being distributed to potential vendors for quote.

It was the opinion of Chemetics that they would not produce liquid SO₂ from our regenerator offgas stream. Chemetics uses a standard cryogenic process to produce liquid SO₂. The gas stream would have been incinerated to oxidize combustibles to either SO₂ or CO₂. The gas stream would next be dried by a dilute acid stream before being compressed and chilled to produce liquid SO₂. The main problem producing liquid SO₂ from the regenerator offgas using

this process is that after incinerating and drying, the gas stream would consist of essentially equal parts CO_2 and SO_2 . Upon compression and liquification, the SO_2 would liquify with unacceptable levels of CO_2 . An additional detriment to using the cryogenic process to produce liquid SO_2 from the regenerator offgas is the gas stream entering the condenser contains 70% inerts, 85% inerts if CO_2 is included, thus the SO_2 yield is poor for the amount of gas which must be compressed and chilled to liquify the SO_2 gas.

However, there are mature absorption technologies which separate the sulfur dioxide from the SO_2 containing gas by reversible absorption into a liquid. Upon desorption of the SO_2 from the stripping liquor, an SO_2 rich gas stream is produced which is easily liquified.⁵ These technologies would eliminate or greatly reduce the problems encountered using only a cryogenic process to produce liquid SO_2 from the regenerator offgas. Several companies have been identified as potential vendors of a liquid SO_2 plant and will be sent bid specification packages.

Once technical feasibility has been established for all three sulfur by-product options and competing bids are received, competitive economics for producing each by-product will be determined.

3.5.2.4 HCl

A potential benefit of the NOXSO process beyond the removal of SO_2 and NO_x from flue gas is the removal of HCl. Measurements made by an independent laboratory at the end of the 5 MW pilot plant test program indicated that NOXSO sorbent adsorbs 95-97% of the HCl in flue gas. However, it is not clear what the fate of the chloride species is after adsorption. Only about 55-70% of the adsorbed chloride was detected in the sorbent heater off gas. Based on solid phase chloride measurements made during the 10,000 hours of pilot plant operation, it is known that chlorides do not accumulate on the sorbent. **Figure 3-13** shows the sorbent chlorine content over 5000 hours of pilot plant operation. Therefore, all the chlorides that are adsorbed must later be desorbed somewhere in the process. Several possibilities exist and these are discussed below.

The first possibility is that the HCl removal efficiency is not as high as the measurements indicate. This would be possible if HCl in the flue gas undergoes a gas phase reaction in the NOXSO adsorber and is converted to a chloride compound that is not detectable with the analyzers used for these measurements. A second possibility is that the chlorides desorb from the heater as something other than HCl which again is not detectable by the analyzers used. Measurements made by NOXSO detected only trace amounts of chloride in the regenerator offgas which is the only other location chlorides could exit from the system. Because of this uncertainty over the fate of chlorides within the NOXSO process, a test program has been initiated at the research center to further investigate all possibilities.

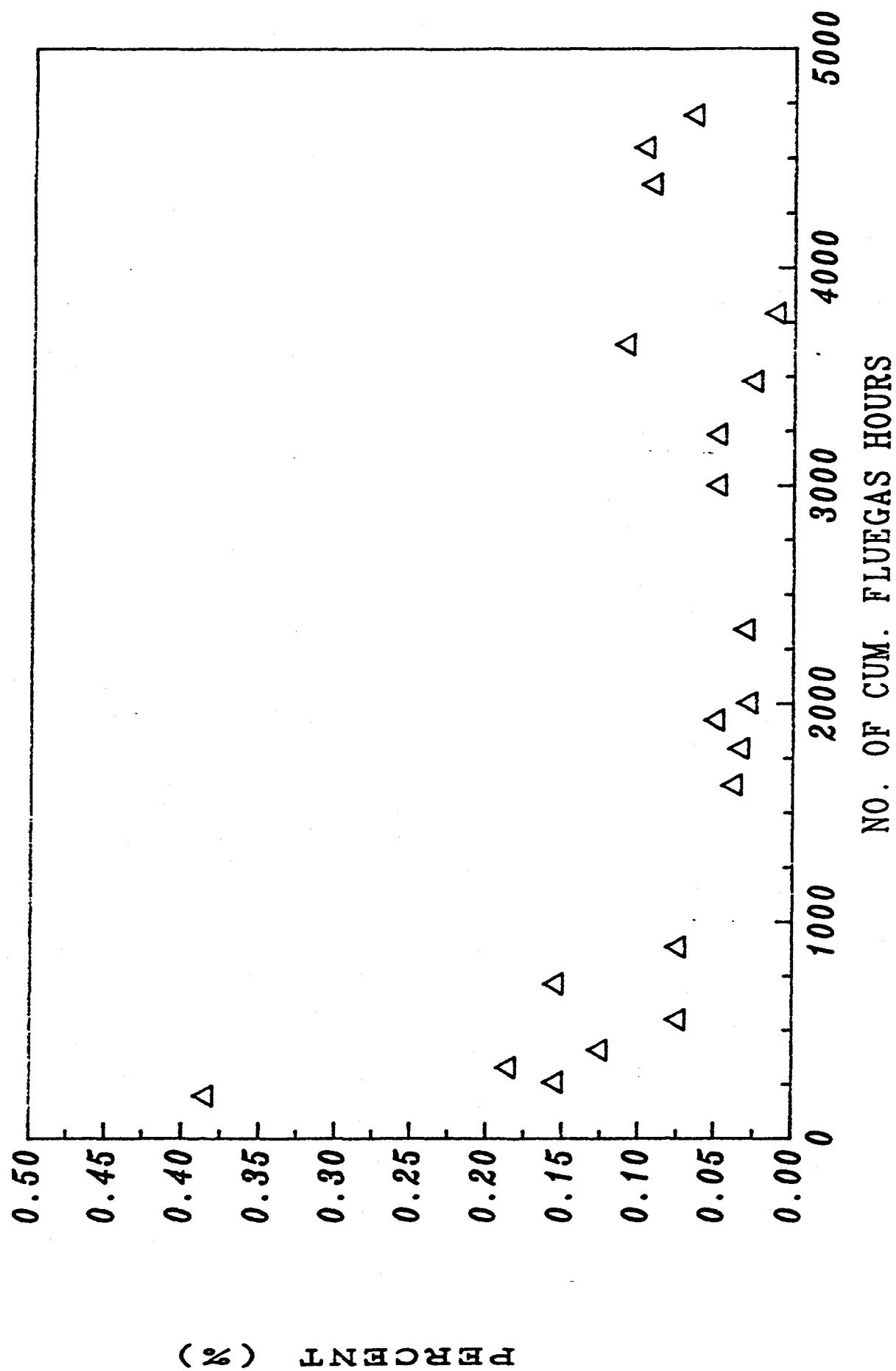


Figure 3-13. Chlorine Content
POC Sorbent

A Milton Roy Prospec 2000 mass spectrometer will be used to analyze offgas streams during both adsorption and sorbent heating in an attempt to identify all chloride compounds present and to close the chloride material balance. In addition, means of achieving the ultimate disposal of the chlorides is under investigation. For example, assuming all the adsorbed chlorides are desorbed in the sorbent heater, two possibilities exist for removing them from the sorbent heater offgas. The first is to further cool the sorbent heater offgas (after the NO_x recycle cooler) to below the HCl acid dewpoint using a water spray. The condensed acid would then be neutralized. In addition to condensing HCl, some nitric acid (HNO₃) would be condensed also. The HCl and HNO₃ dewpoints can be calculated as follows⁶:

For HCl,

$$1,000/T_{dp} = 3.7368 - 0.1591 \ln (P_{H_2O}) - 0.0326 \ln (P_{HCl}) + 0.00269 \ln (P_{H_2O})$$

and, for HNO₃,

$$1000/T_{DP} = 3.6614 - 0.1446 \ln (P_{H_2O}) - 0.0827 \ln (P_{HNO_3}) + 0.00756 \ln (P_{H_2O})$$

Where

$$\begin{aligned} T_{DP} &= \text{acid dewpoint, deg K} \\ P_{H_2O} &= \text{partial pressure of water, mm Hg} \\ P_{HCl, HNO_3} &= \text{partial pressure of acid, mmHg.} \end{aligned}$$

A second means of ultimate chloride disposal would be to concentrate them, adsorb them on a solid material, and then dispose of the solids. Laboratory tests will identify the rate of chloride desorption over a range of temperatures. A sorbent preheater could be used to desorb the chlorides at a lower temperature (less than the 1150°F required for sulfur regeneration) in a small volume gas stream. This gas stream would be treated to remove the chlorides and then become part of the NO_x recycle stream.

3.5.2.5 L-valve Design

The NOXSO process requires that sorbent be transported between six process vessels (adsorber, sorbent heater, regenerator, steam treater, sorbent cooler and surge tank). Two of the sorbent transfers are done by gravity--regenerator to steam treater and sorbent cooler to surge tank. At the pilot plant, one transfer is accomplished by means of a dense phase pneumatic lift--adsorber to sorbent heater. The other three sorbent transfers are accomplished by means of a non-mechanical valve--sorbent heater to regenerator, steam treater to sorbent cooler, and surge tank to adsorber. The non-mechanical valve configuration used at the NOXSO pilot plant is shown in **Figure 3-14**. This "J" configuration was successful in transporting sorbent and isolating vessels at the pilot plant. However, this particular non-mechanical valve

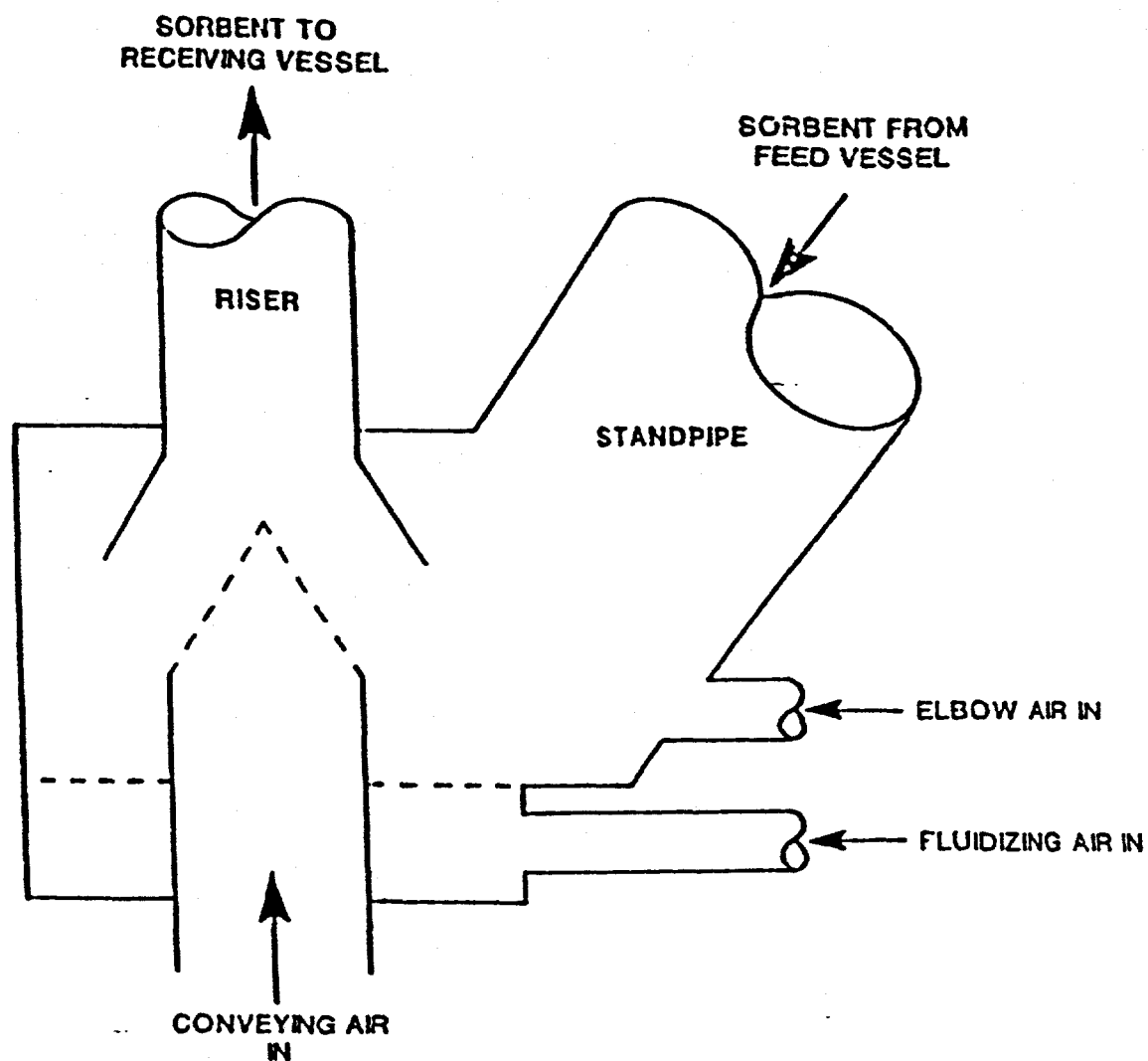


Figure 3-14. J-Valve: Box Detail

configuration also had some shortcomings that are eliminated by changing to an "L" configuration as shown in **Figure 3-15**.

The two drawbacks of the J-valve are 1) it requires three independent gas sources for operation, and 2) it contains several perforated plates that are prone to plugging with sorbent. The L-valve contains no perforated plates and has only two gas sources for operation. Because of these advantages, the L-valve was first tested at the NOXSO Research and Development Center and then a prototype was installed and operated at the pilot plant.

The plexiglass L-valve used in laboratory studies is shown schematically in **Figure 3-16**. The downcomer and riser lengths closely approximate the pilot plant J-valve used to transport sorbent from the surge tank to the adsorber. The effects of conveying gas flow rate, elbow gas flow rate, and feed tank pressure on solids flow rate were studied. A discussion of the test results and a comparison to theoretical predictions is given below.

The pressure profile in the L-valve is critical in determining its performance. Therefore, riser pressure drop was measured over a range of operating conditions and the results were compared to theoretical calculations. The riser pressure drop was calculated as follows:

$$\Delta P_{riser} = \frac{\rho g l \sin \Theta}{g_c} + \frac{U_s G_s}{g_c} + \Delta P_{frict}$$

where

$$\Delta P_{frict} = \frac{2 f_g \rho_g U_g^2 l}{g_c d_t} + \frac{2 f_s G_s U_s l}{g_c d_t}$$

- d_t = riser diameter
- f_g = gas phase friction factor
- f_s = solid phase friction factor.
- G_s = solids mass flux
- g = gravitational constant
- g_c = conversion factor
- l = riser length
- U_g = gas velocity
- U_s = solids velocity
- ρ = mean density of solids and gas in riser
- ρ_g = gas density
- Θ = angle of inclination of riser

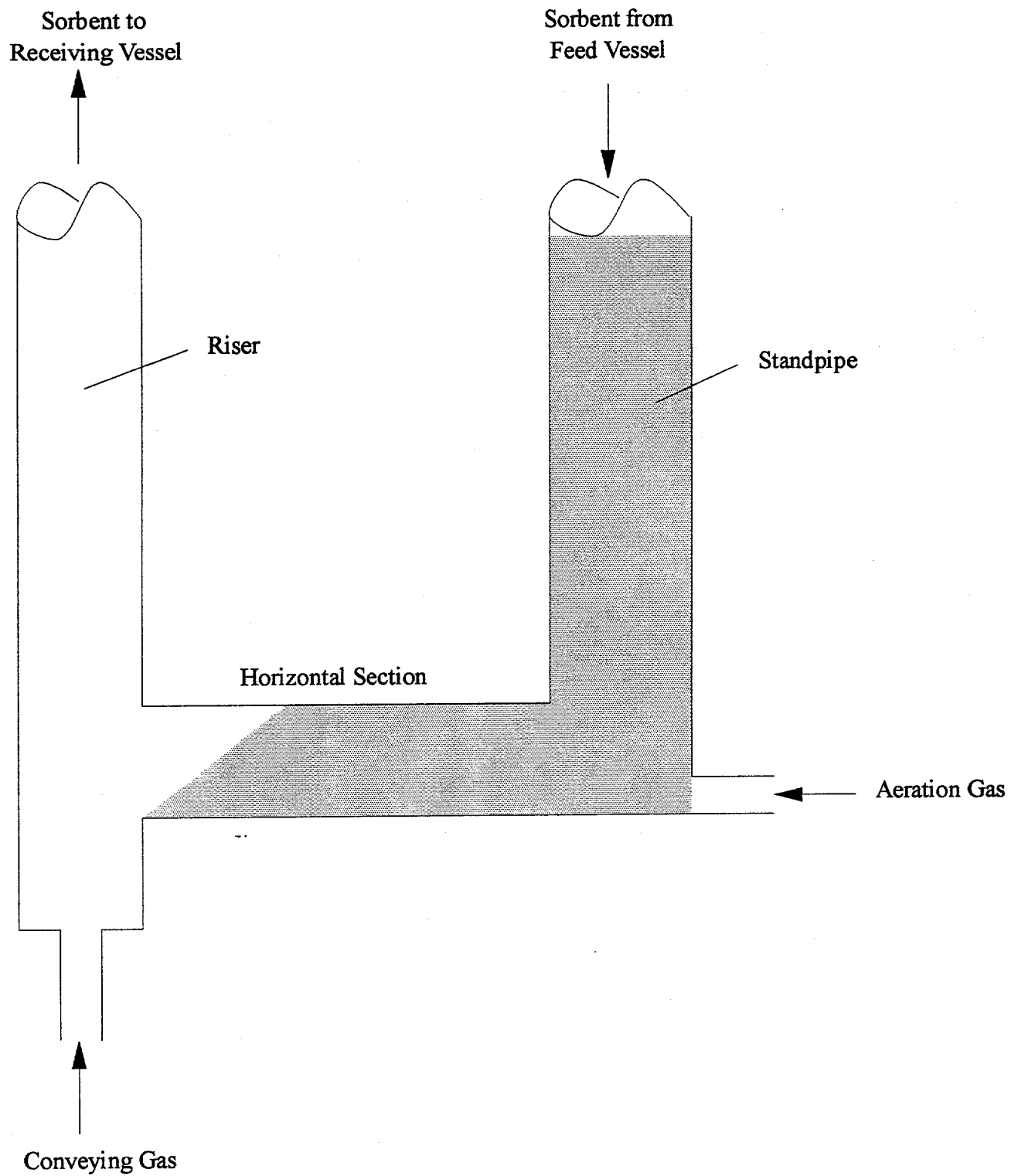
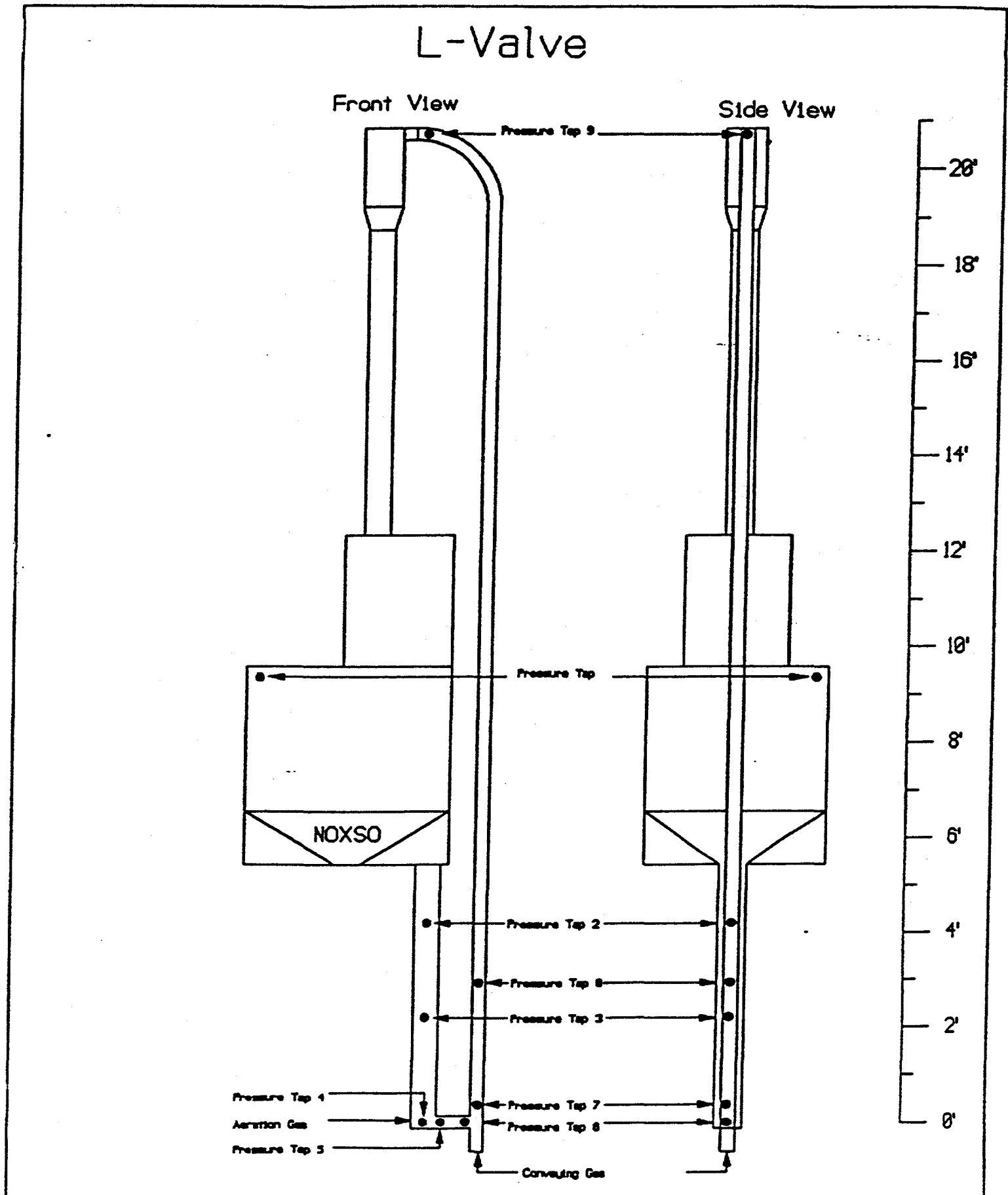


Figure 3-15. L-Valve: Detail

Figure 3-16. L-Valve



The gas phase and solid phase friction factors can be calculated using the following equations:

$$f_g = 0.0791 Re_t^{-0.25} \quad \text{for } 3 \times 10^3 < Re_t < 10^5$$

and

$$f_s = 0.05/U_s \quad (U_s \text{ in m/s}).$$

Table 3-9 contains the gas flow rates, feed tank pressure, measured riser pressure drop and calculated riser pressure drop over a range of operating conditions. At low gas and solids flows the agreement between measured and calculated pressure drop is poor. At higher gas and solids flows, the agreement is good. The reason for this is illustrated in **Figure 3-17** where the measured riser pressure drop is plotted against solids flow rate. After an initial rapid rise in pressure drop with increasing solids flow, the measured pressure drop levels off and even decreases as solids flow continues to increase. This is an unexpected phenomena and will be investigated further in the laboratory. The design equation predicts that as gas and solids flow continue to increase, the riser pressure drop should also continue to increase.

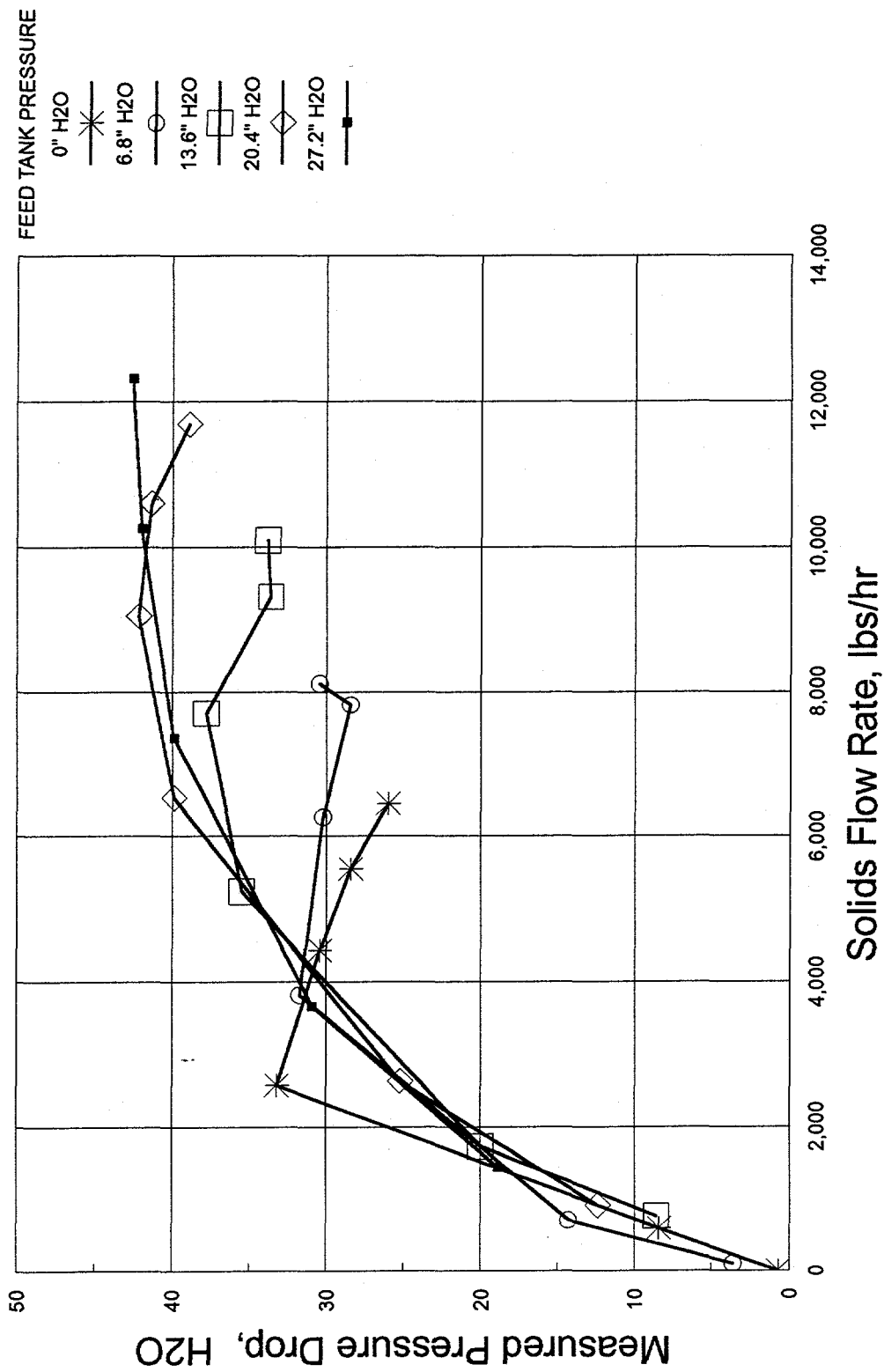


Figure 3-17. Measured Pressure Drop vs. Solids Flow Rate

Table 3-9. L-Valve Operating Conditions and Riser Pressure Drop						
Test#	Feed Tank ("H ₂ O)	Conveying Gas (scfm)	Elbow Gas (scfm)	Solids Flow (lbs/h)	Measured dP ("H ₂ O)	Calculated dP ("H ₂ O)
1	0	50	0	0	0.7	0.5
2	0	50	10	600	8.4	2.7
3	1.4	50	20	2582	33.2	9.6
4	1.4	50	30	4440	30.4	16.1
5	1.4	50	40	5557	28.4	20.3
6	0.7	50	50	6448	26.0	24.0
7	6.8	50	0	100	3.6	1.0
8	6.8	50	10	716	14.3	3.1
9	6.8	50	20	3828	31.7	13.9
10	6.8	50	30	6267	30.2	22.4
11	6.8	50	40	7823	28.4	28.1
12	6.8	50	50	8117	30.4	29.7
13	13.6	50	0	764	8.6	3.3
14	13.6	50	10	1737	20.0	6.7
15	13.6	50	2	5251	35.5	18.9
16	13.6	50	0	7703	37.8	27.4
17	13.6	50	30	9315	33.6	33.5
18	13.6	50	40	10090	33.8	37.0
19	20.4	50	50	912	12.4	3.9
20	20.4	50	0	2642	25.2	9.9
21	20.4	50	10	6530	39.9	23.6
22	20.4	50	20	9048	42.2	32.2
23	20.4	50	30	10601	41.4	38.2
24	20.4	50	40	11689	38.9	42.9
25	27.2	50	50	1451	18.9	5.8
26	27.2	50	50	3675	13.6	13.6
27	27.2	50	50	7358	20.7	26.7
28	27.2	50	50	10249	36.5	36.5
29	27.2	50	50	12324	44.6	44.6

An important feature of an L-valve is the amount of lift that can be achieved. Lift is defined as the difference in elevation between the downcomer inlet and the riser outlet. This lift is dependent on the relative lengths of the downcomer and riser as well as the absolute pressures in the feed vessel and discharge vessel. The low profile plant arrangement discussed previously is predicated on either high temperature dense phase transport systems or alternatively L-valves that can achieve a substantial lift. To test the ability of an L-valve to achieve a high lift, the dense phase transport system at the pilot plant was replaced by an L-valve. In this case, the downcomer had a vertical dimension of just over 22 feet and the riser was 85 feet. The feed tank pressure was about 11" H₂O higher than the discharge vessel pressure. The L-valve was operated successfully over a range of solids flows from 5,000 to 12,000 lbs/h. As in the lab tests, the measured riser pressure drop was higher than the predicted value: ~100" H₂O compared to ~35" H₂O. Again, additional laboratory tests will be conducted to provide insight into this discrepancy.

However, the pilot plant high lift L-valve test did verify that under certain conditions sorbent can be transferred at suitable rates through a 60 plus foot lift. Based on these results, additional laboratory tests will be performed to develop predictive equations for L-valve performance. Ultimately, it will be determined whether or not L-valves can provide the necessary lift in the low profile plant design. A substantial capital cost savings would be achieved by elimination of the high temperature dense phase lifts from that design.

3.6 Plant Characterization

Plant characterization activities are on hold until a new host site is identified.

3.7 Site Survey/Geotechnical Investigation

Site survey/geotechnical investigation activities are on hold until a new host site is identified.

3.8 Permitting

Permitting activities are on hold until a new host site is identified.

4.0 PLANS FOR NEXT QUARTER

The main priority for next quarter is the evaluation and selection of a host site for the project. It is essential that a technically acceptable site be selected so the process can be properly demonstrated.

Immediately upon identification of the host site, work will begin to modify the EIV with information specific to the new site. It is critical to satisfy the NEPA requirements as soon as possible to prevent this delaying the project.

Additional information regarding the low profile plant design will be collected and evaluated. The scaled up POC and low profile designs will be compared from the standpoint of cost and technical risk and the basic design for the commercial design will be selected. Work will continue with potential vendors to identify qualified companies to provide plants for production of both sulfuric acid and liquid SO_2 . Having this information available will enable completion of the study of the sulfur by-product options and enable timely decisions regarding the sulfur by-product of choice for specific sites.

The need to perform additional NO_x destruction studies will be evaluated based on the boiler type for the new host site. However, since a significant data base of NO_x destruction efficiency versus boiler type currently exists, it is unlikely that additional experimental work will be required.

The pilot plant operations were completed on July 30, 1993. Final disposition of the plant will be decided this quarter. The four options being considered include: moving to a new host site, installing a coal combustor at the Toronto site, constructing a closed loop flue gas circuit with injection of SO_2 , NO_x , HCl , etc., and demolition. Based on the ultimate fate of the plant, a decision regarding material inspection, nondestructive or destructive, will be made.

The process studies which are ongoing in support of the commercial plant design will continue. These include modeling of the adsorption and regeneration kinetics to provide insight to optimize both process steps. The computer process simulator will be continually updated based on new and better data describing the process physics and chemistry. The sulfur by-product options study should be completed this quarter providing all the technical and marketing information required to select the best sulfur by-product option for a specific site. Additional laboratory work will be conducted to determine the ultimate fate of the HCl in the flue gas. If appropriate, techniques to remove the HCl will be evaluated to enhance the overall benefit of the NOXSO process by adding HCl to the list of acid rain precursor gases removed. Additional laboratory work and theoretical studies will be conducted to provide accurate L-valve design equations to allow confidence in scaling up to commercial size plants. Laboratory work will be conducted to characterize the size for the attrited sorbent to provide the design information necessary to specify separation equipment for the commercial plant.

As soon as a new host site is identified, activities to collect specific plant information, collect site and geotechnical information, and identify necessary permits will be initiated.

5.0 REFERENCES

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APPENDIX I

ASME BOILER AND PRESSURE VESSEL CODE
SECTION VIII DIVISION 1
APPENDIX I
VESSELS WITH AN OBOUND CROSS SECTION

ADSORBER
GENERAL DATA

PRESSURE P= 4.00 PSI MAX. TEMP. = 400.00 DEG F
RADIUS R= 72.00 IN 1/2 SIDE L2= 48.00 IN

SHELL MATERIAL SA-285 GR C

STRESS S= 13800.00 PSI STRESS SY= 25700.00 PSI
THICKNESS t1= 0.50 IN THICKNESS t2= 1.00 IN

PARA 13-11 REINFORCED VESSEL

STANDARD I BEAM
8.00 IN X 18.40 lb/ft
MATERIAL SA-36

STRESS S= 14500.00 PSI STRESS SY= 30800.00 PSI
AREA A1= 5.34 SQ IN I1= 56.90 IN^4

PARA 13-8 (d)
PITCH $p = t \cdot \sqrt{S \cdot J / P} =$ 41.53 IN J= 2.00
SET p= 24.00 IN

COMPOSITE SECTION

AREA AP=p*t= 12.00 SQ IN A=A1+AP= 17.34 SQ IN

TO FIND NUETRAL AXIS SUM MOMENTS ABOUT INSIDE EDGE OF PLATE

$c_i = (AP \cdot t/2 + A1 \cdot (W1 + t)) / A =$ 1.56 IN
 $c_o = c_i - (W1 + t) =$ -6.94 IN

USE PARALLEL AXIS THEOREM TO FIND I FOR THE SECTION

$I_p = p \cdot t^3 / 12 =$ 0.25 IN^4
 $d_p = c_i - t/2 =$ 1.31 IN $d_1 = -c_o - W1/2 =$ 2.94 IN

$I_{11} = (I_p + A_p \cdot d_p^2) + (I_1 + A1 \cdot d_1^2) =$ 123.90 IN^4

PARA. 13-11 (e)

(1) MEMBRANE STRESS

$(S_m)B = PR_p / (A1 + p \cdot t) =$ 398.62 PSI
 $(S_m)C = P \cdot (R + L2) \cdot p / (A1 + p \cdot t) =$ 664.36 PSI

(2) BENDING STRESS

$$\begin{aligned} r=R+ci &= 73.56 \text{ IN} & \text{GAMMA1}=L2/r &= 0.65 \\ C2=r^2(2*\text{GAMMA1}^2+3*\text{PI}*\text{GAMMA1}+12) &= 102816.03 \text{ SQ IN} \\ A3=r(2*\text{GAMMA1}+\text{PI}) &= 327.09 \text{ IN} \\ (\text{Sb})B=(P*L2*p*co/(6*I11))*(3*L2-C2/A3) &= 7328.65 \text{ PSI} \\ (\text{Sb})C=(P*L2*p*co/(6*I11))*(3*(L2+2*r)-C2/A3) &= -11660.65 \text{ PSI} \\ (\text{Sb})A=(P*L2*p*co/I11)*(-C2/(6*A3)) &= 13524.28 \text{ PSI} \end{aligned}$$

(3) TOTAL STRESS

$$\begin{aligned} (\text{ST})B &= (\text{Sm})B + (\text{Sb})B = 7727.26 \text{ PSI} \\ (\text{ST})C &= (\text{Sm})C + (\text{Sb})C = -10996.29 \text{ PSI} \\ (\text{ST})A &= (\text{Sm})A + (\text{Sb})A = 13922.89 \text{ PSI} \end{aligned}$$

PARA. 13-4 (b) (2) ALLOWABLE STRESS IS THE LESSER OF

$$\begin{aligned} 1.5*S*E &= 18487.50 \text{ PSI} & E &= 0.85 \\ 2/3*Sy &= 20543.60 \text{ PSI} \end{aligned}$$

VESSEL WEIGHT

$$\begin{aligned} W_{\text{shell}} &= (4*L2*t + \text{PI}*((R+t)^2 - R^2)) * L * \text{ROWst} = 24.05 \text{ tons} \\ W_{\text{reinforce}} &= (4*L2 + 2*\text{PI}*(R+t+W1/2)) * L / p * w1 = 11.35 \text{ tons} \\ \text{TOTAL WEIGHT (excluding endplates)} &= 35.39 \text{ tons} \end{aligned}$$

PARA 13-12 STAYED VESSEL

PARA UG-50 MAXIMUM STRESS IN STAYS

$$1.1*S = 15950.00 \text{ PSI}$$

PARA 13-12 (c) STAY MINIMUM DIAMETER

$$t3 = \text{SQRT}(4*P*L2*p/(\text{PI}*S)) = 0.61 \text{ IN} \quad \text{SET } t3 = 0.75 \text{ IN}$$

PARA 13-12 (c) (1) MEMBRANE STRESS

$$\begin{aligned} (\text{Sm})B &= P*R/t1 = 576.00 \text{ PSI} & (\text{Sm})B &= P*R/t2 = 288.00 \text{ PSI} \\ (\text{Sm})C &= P*(R+L2)/t = 960.00 \text{ PSI} \end{aligned}$$

STAY BARS

$$\text{Sm} = 4*P*L2*p/(\text{PI}*t3^2) = 10430.38 \text{ PSI}$$

PARA 13-12 (c) (2) BENDING STRESS

$$\begin{aligned} \text{GAMMA} &= L2/R = 0.67 & \text{ALPHA2} &= I2/I1 = 8.00 \\ A &= R*(2*\text{GAMMA} + \text{PI}*\text{ALPHA2}) = 1905.56 \text{ IN} \\ C1 &= R^2*(2*\text{GAMMA}^2 + 3*\text{GAMMA}*\text{PI}*\text{ALPHA2} + 12*\text{ALPHA}) = 762848.26 \text{ SQ IN} \\ B &= R^2*(\text{GAMMA}^2 + \text{PI}*\text{GAMMA}*\text{ALPHA2} + 2*\text{ALPHA2}) = 172106.75 \text{ SQ IN} \\ D1 &= R^3*(\text{GAMMA}^3 + 2*\text{PI}*\text{GAMMA}^2*\text{ALPHA2} + 12*\text{GAMMA}*\text{ALPHA2} + 2*\text{PI}*\text{ALPHA2}) \\ &D1 = 5.11\text{E}+07 \text{ IN}^3 \\ E1 &= R^3*(4*\text{GAMMA}^3 + 6*\text{PI}*\text{GAMMA}^2*\text{ALPHA2} + 24*\text{GAMMA}*\text{ALPHA2} + 3*\text{PI}*\text{ALPHA2}) \\ &E1 = 1.01\text{E}+08 \text{ IN}^3 \end{aligned}$$

$$\begin{aligned}
 F &= (3 \cdot A \cdot D1 - 2 \cdot B \cdot C1) / (A \cdot E1 - 6 \cdot B^2) = 1.91 \\
 I1 &= b \cdot t1^3 / 12 = 0.01 \text{ IN}^4 & c1 &= t1 / 2 = 0.25 \text{ IN} \\
 I2 &= b \cdot t2^3 / 12 = 0.08 \text{ IN}^4 & c2 &= t2 / 2 = 0.50 \text{ IN} \\
 (Sb)B &= P \cdot L2 \cdot c / (2 \cdot I1 \cdot A) \cdot (F \cdot (B - A \cdot L2) - C1 / 3 + A \cdot L2) = -10535.44 \text{ PSI} \\
 (Sb)C &= P \cdot L2 \cdot c / (2 \cdot I1 \cdot A) \cdot (F \cdot B - A \cdot L2 - A \cdot R) + A \cdot (L2 + 2 \cdot R) - C1 / 3 \\
 & & (Sb)C &= 4229.69 \text{ PSI} \\
 (Sb)A &= P \cdot L2 \cdot c / (2 \cdot I2 \cdot A) \cdot (B \cdot F - C1 / 3) = 90213.14 \text{ PSI} \\
 (Sb)B &= P \cdot L2 \cdot c / (2 \cdot I2 \cdot A) \cdot (F \cdot (B - A \cdot L2) - C1 / 3 + A \cdot L2) = -10535.44 \text{ PSI}
 \end{aligned}$$

(3) TOTAL STRESS

SEMICYLINDRICAL SECTION

$$\begin{aligned}
 (ST)B &= (Sm)B + (Sb)B = 11111.44 \text{ PSI} \\
 (ST)C &= (Sm)C + (Sb)C = 5189.69 \text{ PSI}
 \end{aligned}$$

SIDE PLATES

$$\begin{aligned}
 (ST)A &= (Sm)A + (Sb)A = 90501.14 \text{ PSI} \\
 (ST)B &= (Sm)B + (Sb)B = 10823.44 \text{ PSI}
 \end{aligned}$$

PARA.13-4 (b) (2) ALLOWABLE STRESS IS

$$1.5 \cdot S \cdot E = 17595.00 \text{ PSI} \quad E = 0.85$$

VESSEL WEIGHT

$$\begin{aligned}
 W_{\text{shell}} &= (4 \cdot L2 \cdot t2 + \pi \cdot ((R + t1)^2 - R^2)) \cdot L \cdot \text{ROWst} = 31.19 \text{ tons} \\
 W_{\text{stays}} &= (\pi / 4 \cdot t3^2 \cdot 2 \cdot R \cdot L / p) \cdot \text{ROWst} = 0.20 \text{ tons} \\
 \text{TOTAL WEIGHT (excluding endplates)} &= 31.39 \text{ tons}
 \end{aligned}$$

UG-34 UNSTAYED FLATHEADS AND COVERS

HEAD MATERIAL SA 515 GR 70

$$\text{STRESS } S = 17500.00 \text{ PSI} \quad \text{TEMPERATURE} = 400.00 \text{ DEG F}$$

PARA UG-34 (c) (3)

$$\begin{aligned}
 Z &= 3.4 - 2.4 \cdot d / D = 1.96 & \text{PARA 13-4 (f) } C &= 0.20 \\
 & & E &= 0.85
 \end{aligned}$$

$$t = d \cdot \text{SQRT}(Z \cdot C \cdot P / (S \cdot E)) = 1.48 \quad \text{SET } t = 1.63$$

$$W_{\text{head}} = (\pi \cdot R^2 + 2 \cdot L2 \cdot 2 \cdot R) \cdot t \cdot \text{ROWst} = 6.90 \text{ tons}$$

$$\begin{aligned}
 \text{TOTAL WEIGHT OF REINFORCED VESSEL} &= 49.19 \text{ tons} \\
 \text{TOTAL WEIGHT OF STAYED VESSEL} &= 45.19 \text{ tons}
 \end{aligned}$$